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PROCESS COSTS AND ECONOMICS OF PYRITE-COAL UTILIZATION

BECHTEL CORPORATION
DECEMBER 1968

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PROCESS COSTS AND ECONOMICS OF PYRITE-COAL UTILIZATION

A Report to
Process Control Engineering Division
National Air Pollution Control Administration
Consumer Protection and Environmental Health Service
Public Health Service
U. S. Department of Health, Education, and Welfare
Contract No. PH-86-27-224

BECHTEL CORPORATION
DECEMBER 1968

FOREWORD

Bechtel Corporation, assisted in certain areas by Stanford Research Institute, has conducted a study of the process costs and economics of pyrite-coal utilization. This document reports the results of the Phase I and II study. It was prepared and is submitted in accordance with Contract No. PH 86-67-224, dated June 19, 1967, and amendments thereto.

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Section 1

SUMMARY AND CONCLUSIONS

Industries that burn coal emit significant amounts of air pollutants, a major one being sulfur dioxide. One method of reducing sulfur oxides emission is to remove some or most of the pyritic sulfur from high-sulfur coals before combustion. This coal preparation step produces clean, low-sulfur coal, but it also yields large quantities of sulfur-containing refuse.

The cost of coal preparation has limited its use. However, if a process were developed that would economically recover heat and sulfur values from coal cleaning refuse, this could offset the cost of deep cleaning. Additionally, such development would result in an important source of low-sulfur coal. Potential consumers would include a wide range of industries, other than utilities, in areas that have strict legislation on coal sulfur content.

The study, for the National Air Pollution Control Administration (NAPCA), reviews and compares suitable processes for recovering byproducts and fuel value from coal refuse. It was conducted in two phases. In the first, a number of methods for recovering products were investigated. In phase II, the more promising methods were investigated in greater detail. Results are generally positive and indicate that the selected process could economically utilize coal refuse under certain conditions. It also shows that process improvements are possible which could further help the overall feasibility. The Bechtel study is one part of a broad program by NAPCA to develop pyrite-coal technology. Section 2 of this report presents an overall concept of this program.

REFUSE COMPOSITION

Refuse composition depends upon the composition of the run-of-mine coal and its response to coal preparation techniques. In this study, interest generally concerns bituminous coals indigenous to the eastern United States. Data on typical coals of this type showed sulfur contents ranging from about 1.5 to 5.0 percent and ash contents from about 8 to 25 percent.

Coal preparation techniques remove only the pyritic (FeS_2) sulfur from coal but none of the organic sulfur chemically combined with the coal molecules. In the United States, the percentage of total sulfur attributable to pyrite ranges from 40 to 80 percent.

Generally, coal preparation plants for cleaning steam coals for power plants are not designed for maximum desulfurization. However, techniques exist for removing most of the pyrite and are practiced by some of the steel companies who need metallurgical coke. A maximum sulfur level of 1.0 percent for cleaned coal was set as a criterion in this study. Washability data on coals cleanable to this sulfur level yielded refuse with sulfur contents ranging from about 4 to 25 percent. Ash contents ranged from 20 to 90 percent.

One class of refuse processing methods incorporates adaptations from commercial processes that burn pyrite ore to produce sulfuric acid. These require ore with a minimum sulfur content of about 42 percent. Therefore, if pyrite-coal mixtures are to be used as feed for producing acid in this way, they must be concentrated to higher pyrite grade. Although pyrite concentration was practiced inefficiently on a small scale in the United States during World Wars I and II, no commercial plant exists now. However, many improved mineral dressing techniques have developed in recent years and the technology exists for designing efficient pyrite concentrators.

COMPARISON OF PROCESSES

Two general classes of processes emerge for utilizing coal refuse. One class can feed the entire refuse stream from coal preparation. The other requires a pyrite concentrate feed. Figure 1-1 shows the specific processes that fall into the two classes. Table 1-1 summarizes process characteristics. None of the processes which can use the entire refuse stream has reached commercial status. Processes which feed a pyrite concentrate are commercial but have not been operated with pyrite-coal mixtures. The Combustion-Sulfur Oxides Removal process shows the most favorable economic outlook and its technology appears sound.

In general, processes were selected for review if some major features of their technology appeared adaptable to a pyrite-coal feed. Selection of the most promising process for further study was based on both technology and economics. A few summary remarks follow about specific processes that were considered, but appeared to have less promise than Combustion - Sulfur Oxides Removal. More detailed comments appear in the appropriate sections of the report.

Carbonization - Desulfurization and Solvent Extraction Processes

These two processes are interesting because they do not require a pyrite concentrating step and they produce elemental sulfur. However, both processes are complex and have difficult technical problems to

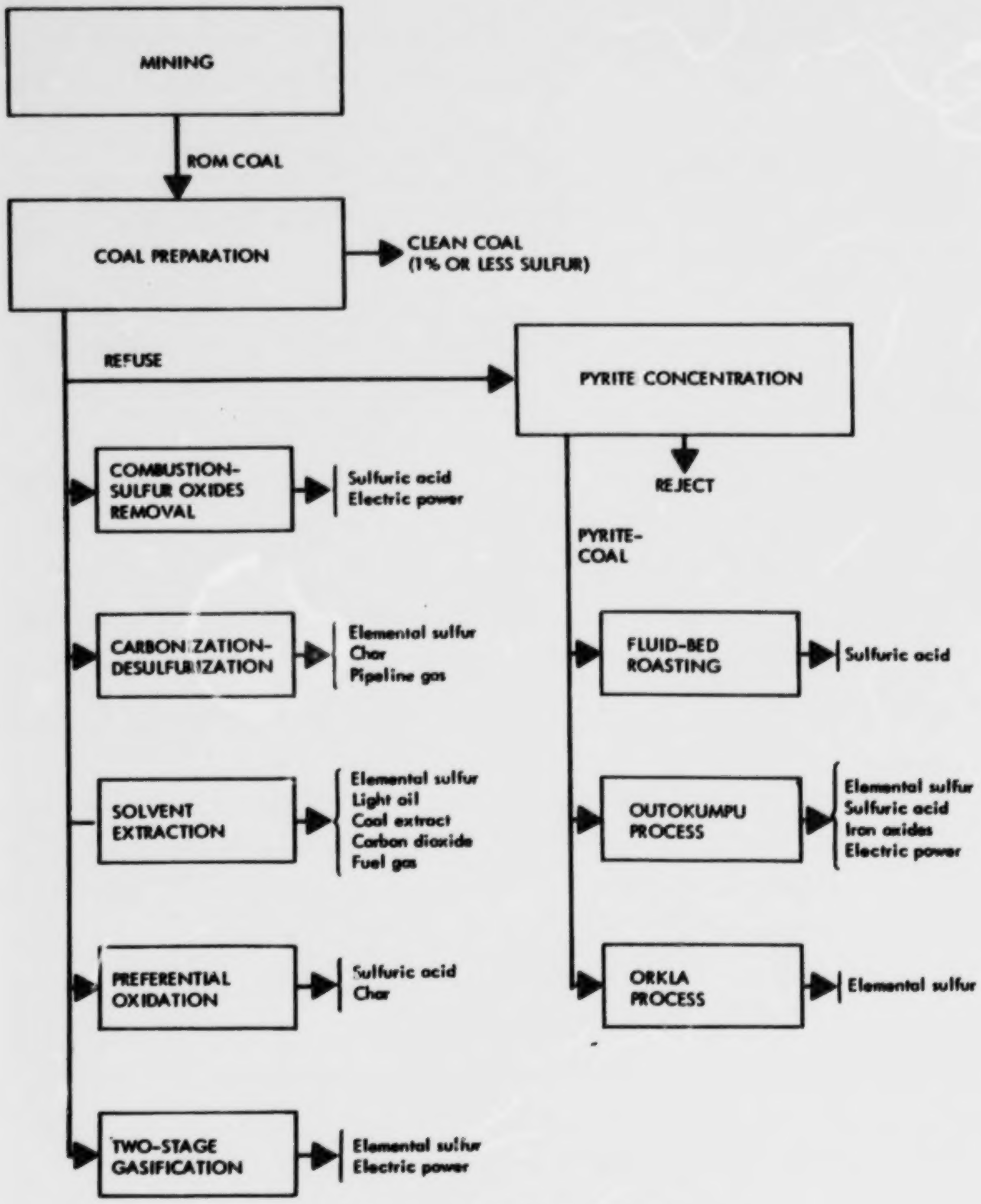
MINING
ROM COAL
COAL PREPARATION
CLEAN COAL (1% OR LESS SULFUR)
REFUSE
PYRITE CONCENTRATION
REJECT
PYRITE-COAL
COMBUSTION-SULFUR OXIDES REMOVAL
Sulfuric acid
Electric power
CARBONIZATION-DESULFURIZATION
Elemental sulfur
Char
Pipeline gas
SOLVENT EXTRACTION
Elemental sulfur
Light oil
Coal extract
Carbon dioxide
Fuel gas
PREFERENTIAL OXIDATION
Sulfuric acid
Char
TWO-STAGE GASIFICATION
Elemental sulfur
Electric power
FLUID-BED ROASTING
Sulfuric acid
OUTOKUMPU PROCESS
Elemental sulfur
Sulfuric acid
Iron oxides
Electric power
ORKLA PROCESS
Elemental sulfur

Figure 1-1. Refuse Processes

Table 1-1

SUMMARY OF PROCESSES FOR UTI

	FEED	PRODUCTS	PROCESS DESCRIPTION
Combustion-Sulfur Oxides Removal	Entire refuse stream	Sulfuric Acid (78%) Electricity	Coal refuse is burned in a ste generator designed for high and sulfur fuel. Sulfur is ox to sulfur dioxide which is cat ically converted to SO_3 follo by recovery of sulfuric acid. Process is in developmental s
Carbonization-Desulfurization	Entire refuse stream	Sulfur Pipeline gas Char	Coal refuse is dried, devolati then desulfurized in a fluid b with gas containing about 50 Sulfur forms H_2S which is r moved by a dolomite accepto Dolomite is regenerated, H_2 is converted to elemental su process gas is converted to gas by a methanation reacti
Solvent Extraction	Entire refuse stream	Sulfur Light oil Coal extract Carbon dioxide Fuel gas	Coal fraction of the refuse is dissolved in solvent derived from the coal itself in the pr sence of hydrogen. Solution filtered to remove ash. Fil is evaporated to recover sol and liquid coal extract is sol ified. The solvent is distill to yield light oil; the heavie fraction is recycled. Proce gas containing H_2S, remove along with solvent vapor, is passed through an amine tre ment to remove the acid ga Acid gas is processed to re cover elemental sulfur. Pr gas is steam reformed to gi H_2S for recycle and CO_2 by

A

Table 1-1

CESSES FOR UTILIZING COAL REFUSE

DESCRIPTION	ADVANTAGES	LIMITATIONS	ECONOMICS
s burned in a steam esigned for high ash uel. Sulfur is oxidized oxide which is catalyt- rted to SO_3 followed of sulfuric acid. developmental status	Does not require pyrite concentration Based on relatively simple, well-known technical principles Products are usually marketable	Sulfuric acid product is slightly discolored and limited to about 78 percent concentration Market radius is limited by shipping cost of acid (usually about 200 miles maximum) Final solution is required to select economical corrosion resistant materials of construction. Pilot plant status	Most promising process
s dried, devolatized, rized in a fluid bed taining about 50% H_2. s H_2S which is re- dolomite acceptor. regenerated, H_2S d to elemental sulfur, is converted to fuel thanation reaction.	Requires no pyrite concentration; uses entire refuse stream Recovers sulfur in elemental form	Complex, multioperation process with many technical problems to solve High-ash char may not be marketable; this product may retain sulfur Pilot plant status	High capital and net operating costs
of the refuse is solvent derived al itself in the pre- drogen. Solution is emove ash. Filtrate ed to recover solvent, oal extract is solid- solvent is distilled t oil; the heavier recycled. Process ng H_2S, removed olvent vapor, is ugh an amine treat- ove the acid gas. processed to re- ntal sulfur. Process reformed to give ycle and CO_2 byproduct.	Requires no pyrite concentration; uses entire refuse stream Produces high quality products	Complex process; it requires further development and solution of technical problems Economics favor very large-scale operation Pilot plant status	Highest capital and net operating costs of processes evaluated

B

Table 1-1 (Continue

	FEED	PRODUCTS	PROCESS DESCRIPTION
Preferential Oxidation	Entire refuse stream	Sulfuric acid	Sulfur is oxidized preferentiall without burning excessive amounts of coal. This is do by carefully regulating the quantity of air and controllin the temperature.
Two-Stage Gasification	Entire refuse stream	Sulfur Electricity	The gasifier is a specially designed vertical cyclone furna with two burner ports, one a the other. Refuse burns at t upper port with limited air t produce an H_2S rich gas and containing FeS. The slag ru down the walls into the flame the lower port where refuse burned with excess air, conv ing all sulfur compounds to S Gases passing up through fur become H_2S-rich. Subseque treatment recovers sulfur a yields 80-100 Btu/cubic foot gas.
Fluid Bed Roasting	Pyrite concentration	Sulfuric acid Iron oxides Electric power	Pyrite concentrate is roasted fluid-bed furnace to produce and a low-sulfur hematitic c cine. SO_2 is converted to ac a conventional contact plant. ore may or may not be recov depending upon quality and m Waste heat can be recovered process or generation of ele city.

1-1 (Continued)

DESCRIPTION	ADVANTAGES	LIMITATIONS	ECONOMICS
ed preferentially g excessive al. This is done egulating the and controlling re.	Simple process scheme with few operational steps	Treatment removes only a small fraction of organic sulfur and not all pyritic sulfur. Status of development is bench-scale with very limited data. Temperature control problem difficult to solve	No data because of process status
a specially de- l cyclone furnace er ports, one above efuse burns at the th limited air to $_2$S rich gas and slag S. The slag runs s into the flame at t where refuse is xcess air, convert- compounds to SO_2. g up through furnace rich. Subsequent covers sulfur and Btu/cubic foot fuel	Simple process with efficient recovery of fuel value Produces valuable elemental sulfur	The process is in the concept stage only, no development work has been done.	No data because of process status
rate is roasted in a ace to produce SO_2 lfur hematitic cal- converted to acid in l contact plant. Iron ay not be recovered on quality and market. n be recovered for neration of electri-	Relatively simple, well-developed technology	Requires pyrite concentration Requires pilot plant development for use with coal refuse Dilution of pyrite with coal causes lower concentration of SO_2 in gas stream. Degree of acid discoloration is unknown.	Low capital cost Net operating cost is encouraging at this stage of development

B

Table 1-1 (Continued

	FEED	PRODUCTS	PROCESS DESCRIPTION
Outokumpu	Pyrite Concentration	Sulfur Sulfuric acid Iron oxide Electricity	Pyrite is flash smelted by com bustion gases which remove labile sulfur for recovery in mental form. Other gaseous compounds are catalytically verted to S. Liquid FeS matt is granulated and roasted to an iron oxide product and SO for conversion to acid.
Orkla	Cupriferous Pyrite	Sulfur	The Orkla process smelts pyr in a specially designed blast furnace to produce elemental sulfur. Iron, copper, and pr cious metals report in the m

A

le 1-1 (Continued)

DESCRIPTION	ADVANTAGES	LIMITATIONS	ECONOMICS
smelted by com- which remove the for recovery in ele- . Other gaseous re catalytically con- Liquid FeS matte d and roasted to give e product and SO_2 on to acid.	Recovers about half of the sulfur in elemental form Process developed to commercial scale.	Ash content of coal refuse would cause technical problems Economy would suffer because iron sinter would not be marketable. Requires very large-scale operation	Not as favorable as conventional fluid-bed process
ocess smelts pyrite ly designed blast roduce elemental n, copper, and pre- report in the matte.	Produces elemental sulfur Process developed to commercial scale	Limited to a lump pyrite feed Modification of technology for feeding coal refuse would be difficult.	No published data It is generally believed to be economic only with a copper product.

B

solve. These problems would be more acute with coal-pyrites feed than with (run-of-mine) coal. Preliminary economics are not promising.

Outokumpu and Orkla Processes

These two processes are interesting because they produce elemental sulfur. Certain aspects of the chemistry and technology are of interest but the processes would require major modifications to adapt them to pyrite-coal utilization. Economics are not encouraging: Outokumpu feasibility depends upon very large-scale operation and producing a marketable iron oxide (not likely with pyrite-coal); Orkla economics depend on marketing a copper byproduct. Both processes require a pyrite concentrating step.

Preferential Oxidation

Not enough experimental data has been published on this concept for development of a process design or cost estimates.

Fluid - Bed Roasting

This technique is of considerable interest because it involves relatively simple, well-developed technology with modest capital investment. Bechtel did not pursue this process because A. D. Little and Dorr-Oliver are studying it under separate contract to the Public Health Service. Preliminary economics are encouraging but appear slightly less favorable than Combustion-Sulfur Oxides Removal. However, development work may show that this process is an even more serious competitor.

TWO-STAGE GASIFICATION

This process is briefly summarized in Table 1-1 and described in more detail in Section 8. At this time it is strictly in the concept stage. Work shows that gasification of refuse is technically feasible in a specially designed furnace. Recovery of elemental sulfur and efficient use of the fuel value of the refuse are incentives for further consideration of this scheme.

Coal refuse imposes special problems during gasification due to its high ash and sulfur content. The two-stage concept is an adaptation to accommodate a refuse feed material. Some of its specific advantages over single-stage gasification are stated below.

- The standard gas producer leaves powdery ashes containing perhaps as much as 15-percent unburned carbon. The two-stage furnace gasifies nearly all the carbon and discharges most of the ash as a slag which flows from a tap in the furnace bottom.
- The standard single-stage operation produces a gas that contains the organic sulfur from the coal but only the labile sulfur from the pyrite. The sulfur remaining as ferrous sulfide is lost in the ash. The two-stage furnace gasifies nearly all the pyritic sulfur.

COMBUSTION-SULFUR OXIDES REMOVAL

Table 1-1 briefly defines this process and summarizes its characteristics. The technical aspects can be classified into two categories: those concerned with steam generation and those concerned with sulfuric acid production. Acid recovery facilities are closely integrated with the boiler, and require some modification of the steam generator system. The major design considerations for the steam generator, however, concern the high ash content of the refuse feed and the physical properties of this ash. The ash fusion temperature and viscosity determine whether it can be removed from the furnace as a molten slag. If this is not possible, most of the ash will carry through the boiler as fly ash, and the required investment in the dust collection system will be substantially increased.

Almost any coal, including refuse with substantial ash, can be successfully burned by pulverized firing. However, if ash characteristics permit, cyclone firing would have several important advantages. Besides alleviating the fly ash problem, cyclone firing reduces investment in fine grinding below that for pulverized refuse firing. An alternative to either cyclone or pulverized feed steam generators is fluid-bed combustion, a new firing technique in the development stage. Its goal is to decrease the size and cost of boilers. Fluid-bed firing could be attractive for coal refuse also because the low excess air required would result in higher sulfur dioxide concentration.

Many processes are under development for removing sulfur oxides from flue gas. The Monsanto process was selected for the economic estimates because of its advanced developmental state and simplicity. A detailed comparison of competing SO_2 removal processes might improve the overall economic outlook. In general, the slow commercial acceptance of these processes concerns economics rather than technology. For most SO_2 concentrations in flue gases, the cost of recovery exceeds the value of product recovered. The high sulfur content of coal refuse may change this situation.

Although the refuse from coal preparation plants is not currently marketed, it has an intrinsic worth due to its sulfur content and fuel value. A method is developed in Section 7 to calculate the value of coal refuse (its sales price, FOB coal preparation plant) as a function of numerous variables. This method assumes that coal refuse is bought jointly by the owners of a closely integrated electric utility and sulfuric acid plant. In the evaluation method used, each of these two investors would pay for his share of the refuse value (fuel or sulfur). The owner of the coal preparation plant could use this income to pay for all or part of the cost of coal cleaning.

At this stage of consideration, the variables that affect refuse value are many and have rather wide limits for their values. To give the economic analysis the necessary flexibility, the results are presented in a series of graphs. As a result, it is possible by picking specific conditions to make conclusive economic judgments.

An example may help the reader gain a more concrete idea of the use of the data presented in Section 7. 2. A typical problem is to estimate for specific conditions the value of refuse of known composition. Table 1-2 shows the results of such an estimate for two refuse compositions. Both contain 34. 0 percent ash and 15. 0 percent moisture, but differ in total sulfur concentration.

The values in the Total Refuse Value column represent the estimated value of refuse to the owners of the coal preparation plant. The net values represent that associated with each of the two marketable refuse components — energy (coal) and sulfur. In the last column, the percentage reduction in coal preparation cost possible through refuse sale is shown. A minus sign designates a possible reduction in preparation cost and a plus an increase.

Refuse value depends upon many variables, the most important being refuse power plant capital cost and sulfuric acid sales price.

Table 1-2

VALUE OF REFUSE

Refuse Sulfur Concent. (percent)	Sulfuric Acid Price ($/ton)	Net Refuse Energy Value (cents/10^6 Btu)	Net Refuse Sulfur Value ($/ton sulfur)	Total Refuse Value		Change in Coal Prep. Cost (percent)
				($/ton refuse)	(cents/10^6 Btu)	
6.38	10.00	4.0	-10.24	-0.06	nil	2.5 increase
6.38	15.00	4.0	7.70	1.09	7.3	45.4 decrease
10.62	10.00	4.0	7.18	1.35	9.2	56.2 decrease
10.62	15.00	4.0	25.10	3.25	22.2	135.4 decrease

ASSUMPTIONS

Refuse Power Plant Capacity	300 Mw (net)
Refuse Power Plant Capital Cost*	170 $/kw
Return On Book Value	8.0 percent
Refuse Ash Concentration	34.0 percent
Refuse Moisture	15.0 percent
Sulfur Oxides Recovery Plant Investment (in terms of refuse power plant capacity)	42 $/kw
Discounted Rate of Return	12.5 percent
Coal Preparation Plant Yield	80 percent
Coal Preparation Cost	$0.60/ton clean coal

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*Does not include cost for SO_2 recovery plant.

Other factors affecting refuse value which were investigated are:

- Refuse ash concentration
- Refuse power plant operating and maintenance cost
- Sulfuric acid recovery plant investment
- Rate of return on acid recovery plant

Curves showing the influence of these variables are presented in Section 7.

Section 2

INTRODUCTION

Air pollution is already a serious problem in many parts of the country. While the major sources of air pollution vary with the geographic location and characteristics of the environment and community, it is important to note that electric utilities and other industries that burn coal generate a significant proportion of these pollutants in many of the heavily industrialized and urbanized areas.

Sulfur dioxide is one of several major pollutants resulting from the burning of coal. A decrease in the emission of sulfur dioxide from industrial sources could provide a significant overall reduction in sulfur dioxide levels in some localities. For this reason, the Process Control Engineering Division of the National Air Pollution Control Administration has been investigating methods of reducing or eliminating the sulfur dioxide emitted during the combustion of sulfur-containing coals.

One method of reducing sulfur oxides emission is to remove some or most of the pyritic sulfur from the high-sulfur coal burned by electric utilities. Such coal preparation, or cleaning, produces large quantities of high-sulfur refuse. If some process could utilize this refuse in an economic way, it would offset the cost of desulfurizing by coal preparation techniques. This, then, would establish an important source of low-sulfur coal.

A number of industries, other than utilities, need low-sulfur coal. However, it would not be economic for them individually to build and operate a coal preparation plant for deep cleaning. Such potential consumers may be willing to pay a premium for this fuel, especially in areas where legislation strictly regulates allowable coal sulfur content. A premium value for low-sulfur coal would then further justify the feasibility of the pyrite-coal process.

This study was conducted to determine possible ways to utilize the pyrite-rich refuse from coal preparation and to provide more detailed information on the technical and economic feasibility of such utilization. The Bechtel study is one part of a comprehensive program by NAPCA to develop pyrite-coal technology to commercial feasibility.

2.1 RATIONALE OF THE STUDY

The study reported here was initiated in August 1967 by Bechtel Corporation with Stanford Research Institute as a subcontractor on process economics and marketing aspects. To provide for periodic review, and by the logic of the investigation, the study was divided into three phases. In the first phase, a preliminary evaluation was to be made of the available technology, including potential but incompletely developed processes, and the marketing factors applicable to them. As the culmination of this phase, a process was to be selected for full economic and technologic assessment. In the second phase, extensive analysis was to be made of the market potential, a target area selected, and a conceptual design generated, with cost estimates and process economics which would support a determination of feasibility of the process.

Phase I work made it possible to sort out those processes that are most promising for utilizing pyrite-coal from those not suitable. It further indicated that the most appropriate scheme is to burn the pyrite-coal mixture in a special steam generator and recover sulfur dioxide. However, it became apparent that process technology and economics were not developed to a level which would make all the objectives of Phase II achievable. The effort was accordingly directed to further study of technical and economic feasibility of burning pyrite-coal in steam generators.

At this stage of development, economic aspects appeared the most critical. Consequently, an extensive computer analysis was made using a number of parameters to cover a wide range of interest.

The reasons for the promising economics of pyrite-coal combustion led to the concept of a two-stage gasification scheme. Process calculations were made which show that this method may be worthy of further consideration.

2.2 ORGANIZATION OF THE REPORT

Besides the Summary and Introduction, the report contains five major sections with supporting information gathered into appendixes. Section 3 contains the technological overview of coal preparation and pyrite concentration. It then diagrams how the resulting material is treated by either of two general classes of refuse processes.

Sections 4 and 5, respectively, review the technology available for processing coal refuse and pyrite concentrates. Section 6 is essentially a summary of the tradeoffs with respect to technologic feasibility and economic factors which led to the tentative selection of a process for fuller analytic treatment.

Section 7 presents the results of the economic and technical analysis of the Combustion-Sulfur Oxides Removal process. Several Bechtel computer programs were linked together and modified for use in the economic analysis.

Section 8 presents the concept and process calculations for the Two-Stage Gasification study.

2.3 REFERENCES AND ACKNOWLEDGEMENTS

A considerable body of literature was reviewed, and this is listed in the Bibliography. Additional current technical and economic information was supplied by direct contact with companies having specific expertise. In this respect, we wish to express our appreciation to the following organizations:

- Coal Preparation: Consolidation Coal Co.; Pittsburgh Midway Division of Gulf Oil Corp.; Bituminous Coal Research Incorporated; Heyl & Patterson Incorporated; Bethlehem Steel Corp.; United States Steel Corp.; McNally Pittsburgh Manufacturing Corp.; and Pittsburgh Coal Research Center of the U.S. Bureau of Mines.
- Steam Generation: The Babcock and Wilcox Company; Pope, Evans & Robbins; and Foster Wheeler Corporation.

Section 3

TECHNOLOGICAL OVERVIEW

Sulfur occurs in coal in three forms: organic, sulfides, and sulfate. Organic sulfur is chemically combined with the coal molecules and cannot be removed by physical means. Sulfides occur almost entirely as iron disulfides in two crystalline forms, pyrite and marcasite. The term pyrite is commonly used (and is used herein) to include both forms. Pyrite sulfur comes in many particle sizes, the coarser ones being relatively easy to remove, the finely disseminated ones very difficult. The amount of sulfate sulfur (mostly calcium sulfate) is usually negligible from the standpoint of coal preparation.

Both sulfur concentration and form vary widely between coal seams, mines, and even within mines. The total sulfur content of U. S. coals ranges from about 0.2 to 7.0 percent on a dry basis. Pyrite sulfur content ranges from 40 to 80 percent of total sulfur. Generally, organic sulfur predominates in the low-sulfur coals. However, in some cases organic sulfur may be over 2 percent.

The refuse from cleaning run-of-mine bituminous coal provides the feed material for all the processes considered in this study. Composition of the raw coal and techniques used in coal preparation determine the composition of the refuse and the clean coal product. Therefore, to evaluate the feasibility of refuse processing, it was necessary for Bechtel to review the current technical status and economics of coal preparation.

Primary aims were to answer questions concerning:

- Compositions of refuse that are possible and desirable from the standpoints of technology and economics
- Reasonable desulfurization limits
- Possibilities of achieving a clean coal with a sulfur content of 1.0 percent or less
- The ability of current preparation plants to desulfurize to 1.0 percent sulfur
- The suitability of current preparation plant design for integration with refuse processing
- Design changes needed to achieve a high degree of desulfurization
- Appropriate sizes for coal preparation plants
- Costs of coal preparation

These questions were answered at least to the extent necessary for setting reasonable criteria and bases for economic analysis.

In another recent study for NAPCA, the Paul Weir Company(1) has investigated in some detail the desulfurization of coal by generally known coal preparation methods. The Weir study also considered the feasibility of recovering or concentrating the pyrite or a by-product suitable for sulfuric acid manufacture. The reader is referred to this work for details of the technology and economics of coal preparation. Discussion of this subject herein will be brief, its purpose being to show how some of the major aspects of coal preparation relate to refuse processing.

3.1 COAL PREPARATION

Most coals have to meet certain market requirements as to ash, moisture, and Btu values, with penalty and bonus clauses above or below

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[1] "An Economic Feasibility Study of Coal Desulfurization," for the National Air Pollution Control Administration, Department of Health, Education and Welfare, by Paul Weir Company, Inc., Chicago, Illinois, Oct. 1965, Contract No. PH86-65-29.

set figures. Generally, coal preparation plants for cleaning steam coals for power plants are not designed for maximum desulfurization. The much smaller quantity of coal for metallurgical coke requires lower sulfur limits. As a result, certain steel companies have developed some of the most efficient plants for desulfurization. Recent legislation on air pollution has brought the maximum allowable sulfur content of power plant coals down to about 1.0 percent for much of the Eastern United States. This sulfur level for cleaned coal has been established as a necessary requirement of the coal preparation plants assumed in the economic estimates.

3.1.1 Commercial Practice

Only pyritic sulfur is potentially removable from coal by current preparation techniques. The raw coal is first crushed to liberate pyrite particles. The size required for liberation varies from 1.5 inches to 200 mesh. Liberation is followed by separation, which can be accomplished in either of two ways: (1) specific-gravity separation of the lighter coal from the heavier pyrite or other mineral matter, or (2) froth flotation, which is effective for certain particle sizes.

Specific-gravity separations are carried out at densities of 1.60, 1.45, and 1.35 by a variety of devices (e.g., Baum jigs, Chance dense media cones, Deister tables, or hydroclones). Deister Concentrator Co. state that their tables can effectively remove free pyrite from fine coal down to 100 mesh at sp. gr. greater than 1.5. If the feed to the table is suitably classified, the separating gravity can be reduced to 1.4. Selection of the particular type of equipment depends on the size of the fraction being processed. Lower separating densities give better selectivity, but at the expense of yield.

Recent research by Bethlehem Steel shows that froth flotation, in its current state of development, has potential for removing high specific gravity impurities from +100 mesh coal. It may be the best separation method now available for the 14 x 28 mesh range. With current practice, froth flotation has been ineffective in processing -100 mesh coal. Redesign of flotation apparatus to produce smaller bubbles may extend its use to the -100 mesh range.

The Paul Weir report describes six case studies of existing preparation plants washing high-sulfur coals in Illinois, western Kentucky, Ohio, and western Pennsylvania. Plant facilities are rated as better than average. In these plants, the coal was not crushed finer than 1.5 inches and the reporting density was 1.60. The raw coals ranged in sulfur content from 2.66 to 5.49 percent, with organic sulfur varying from 0.95 to 2.20 percent. None of the plants was specifically designed to achieve maximum desulfurization. The flow sheets were relatively simple, with the number of stages of separation kept to a minimum especially in fine coal processing. One plant used hydroclones and one plant used froth flotation.

3.1.2 Proposed Methods for Maximum Desulfurization

The Paul Weir Co., several steel companies, and various vendors of coal preparation equipment state that techniques exist for producing low-sulfur coal. The degree of success is of course limited by the washability characteristics of the coal as indicated by float-sink tests.

A process design for maximum desulfurization would pass the coal through a series of crushing, grinding, and size-classification steps to produce a number of closely sized fractions. The recently developed DSM(2) sieve bend is highly recommended for fine screening. Pyrite would be removed from these fractions by the method most suitable to

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[2] F. J. Fontein, "The D. S. M. Sieve Bend, New Tool for Wet Screening on Fine Sizes, Application in Coal Washeries," presented at Second International Coal Preparation Congress, Essen, Sept. 20-25, 1954.

the particle size. For example, Baum jigs with a separating density of 1.60 might be used for the coarsest fraction. Finer fractions could then be handled in dense-media cones and hydroclones at 1.45 to 1.35 densities. Finally, froth flotation or a combination of this with hydrocloning would be used for fine particle size fractions.

It appears that the coal preparation industry has adequate processes for cleaning coarse coal (+28 mesh). For fine coal processing (-28 mesh), the techniques, which are hydrocloning and froth flotation, are probably suitable but need to be keyed together more effectively. Recently published data by Bethlehem Steel Corp.[3] show progress in this direction. Their work presents data and suggests techniques for froth flotation and for combining it with hydrocloning.

3.2 PYRITE CONCENTRATION

Recovery of pyrite from coal has been practiced only during the sulfur shortages imposed by World Wars I and II. Even then, coal pyrite in Great Britain and the United States accounted for less than one percent of the sulfur requirements. The lack of interest has been due to poor economics rather than to technical problems. Most of the pyrite concentrating plants operated successfully but eventually closed down due to competition from brimstone or conventional pyrite ore.

Conventional processes that burn pyrites and produce sulfuric acid require ore with a minimum sulfur content of about 42 percent. Total sulfur in the refuse from high-sulfur coals varies from about 4 to 25 percent. Therefore, if this refuse is to be a suitable feed for acid production, it must be concentrated to a higher pyrite grade.

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[3] F. G. Miller, J. M. Podgursky, and R. P. Aikman, "Researchers Find Key to the Mechanism of Coal Flotation and Its Role in Fine-Coal Processing Systems," Mining Engineering, July 1967, p. 165-170.

Pyrite concentration during the wars was conducted without the benefit of current effective mineral dressing techniques. One method consisted of hand-picking lumps or nodules of pyrite or a combination of hand-picking followed by a mechanical concentrating table. Unfortunately there was a very limited supply of lump pyrite. The Mineral Products Co. (subsidiary of Pittsburgh & Midway Coal Mining Co.) designed and operated a concentrating plant which produced a satisfactory grade of pyrite. The design, typical of the early plants, was complicated and could now be simplified by recent technology. Eventually the plant was forced to close down due to competition from other sources of sulfur. By 1940, the British designed an improved pyrite concentrating plant, using several stages of crushing and classifying, effectively combined with Baum jigs and Wilfley tables.

At the present, a modern commercial pyrite concentrating plant does not exist. However, the techniques do exist and such a process could be designed. To achieve a pyrite concentrate of 42 percent sulfur, it would be necessary to re-treat the refuse from coal preparation plants. This would be true of refuse from existing plants or from the improved plants described in paragraph 3.1.2.

The Paul Weir study suggests two methods for re-treating a high-pyrite refuse to produce a commercial grade pyrite suitable for acid manufacture:

Case 1	The run-of-mine coal has been processed by an improved preparation plant designed for maximum desulfurization. In this case, the high-sulfur reject is minus 30 mesh, and a final pyrite concentrate can be prepared by treating on Deister tables.
Case 2	The run-of-mine coal has been processed by a typical existing preparation plant where crushing is only 3 inches x 0. The high-sulfur refuse is crushed to 3/4 inches x 0, passed through mineral jigs to remove larger coal or shale particles, and then crushed to six mesh. A pyrite concentrate is finally separated on coarse Deister tables.

3.3 REFUSE YIELDS AND COMPOSITIONS

Modern coal preparation plants treating high sulfur coals produce a refuse stream that may reach 28 percent pyrite; a possible average may be about 15 percent. Improved plants with finer crushing and efficient fine coal preparation circuits may produce refuse with as high as 50 percent pyrite. From 75 to 100 percent of the sulfur present in refuse is pyritic. The amount of a plant reject varies from 5 to 40 percent of the feed to the plant; an average for American plants is about 15 percent.

Specifications for pyrite concentrates sold to sulfuric acid plants often state a maximum carbon range of 5 to 8 percent and a minimum sulfur limit of 42 percent. The sulfur content is perhaps based on sound economics, but the carbon limit should be examined to see if it is realistic in light of modern technology. The low carbon limits appear to have been set in the early days of pyrite-coal burning.

In the past, most of the sulfuric acid from pyrite-coal was manufactured by the chamber process, and it was argued that carbon dioxide dilution reduced catalyst chamber capacity. It was also believed that carbonaceous material fouled the solid catalyst in contact acid plants. However, one British contact plant operated successfully with pyrite-coal, although the actual limits for carbon were never determined.

The minimum sulfur content as currently specified for acid plants is based on both economics and technology. Lower grade pyrite ores usually cannot compete due to high freight cost per unit of sulfur. If the diluent contains considerable silica, fluxing causes difficulty with cinder handling within the furnace. If the diluent is carbon as in coal pyrite, then at some carbon content the gas SO_2 concentration will fall below the design criteria of the acid plant. Extent of acid discoloration and catalyst fouling are unknown at this time. Acid plants can be designed for ores containing less than 42 percent sulfur and for gas streams containing less than 5 percent SO_2. However, their feasibility must be determined for the specific case in question. See subsection 5.1.2 for further discussion of sulfur and carbon contents.

For use in this study, refuse compositions were taken from washability data on coals from western Pennsylvania, eastern Ohio, and northern West Virginia. Appropriate sections of this study discuss the use of these data in economic calculations.

3.4 PROCESSES AVAILABLE FOR REFUSE TREATMENT

The processing of coal refuse is not a new concept. Technologies exist for producing chemicals, fuels, and other products from coal, and for extracting its impurities. Refuse from desulfurizing coal to one percent would have the properties of high-ash, high-sulfur coal, and could be processed like run-of-mine (ROM) coal. As explained before, this refuse contains significant quantities of iron pyrite, but of too low concentration to be treated as commercial pyrite ore.

Two general classes of processes emerge for treatment of coal refuse to produce valuable by-products or to recover its fuel value efficiently. Figure 1-1 presents a diagrammatic outline of processes of these classes. One set uses the entire refuse stream from coal preparation. The other set requires a pyrite concentrate feed, since they were originally designed for commercial pyrite ore. These particular pyrite processes were selected because they could be adapted to feed pyrite-coal.

3.5 PRODUCTS AND MARKETABILITY

3.5.1 Sulfur Compounds

Sulfur compounds are the most important potential products from the standpoints of quantity and revenue. Nearly all commercial processes that burn pyrite convert the sulfur content to sulfuric acid. The Outokumpu and Fluid-Bed processes could produce a concentrated acid (98% H_2SO_4), but probably with some discoloration due to carbon content. The Combustion-Sulfur Oxides Removal process using the Monsanto design would produce acid of about 78% H_2SO_4. Some discoloration results with this method also. The dark acid would limit its use to fertilizer manufacture or for pickling steel, but this might not be serious since about 50 percent of United States sulfuric acid goes to these uses. The market area for sulfuric acid is generally considered to lie within a 200-mile radius.

Figure 1-1 shows that some processes produce elemental sulfur. This expands the market area tremendously, since elemental sulfur is shipped throughout the world. In general it costs more to produce elemental sulfur than acid from pyrite. If a nearby market for acid exists or could be created, the extra expense of reducing sulfur dioxide to elemental sulfur would not be economic.

3.5.2 Iron Oxide

Hematite (Fe_2O_3) is produced when pyrite is roasted under normal operating conditions. However, further processing is required to convert this calcine to a suitable blast-furnace feed. Only a few commercial processes currently do this (in response to a local market for the oxide). In the case of pyrite-coal, the incentives for producing a salable iron ore would be rather low. Unless the hematite calcines were produced through an expensive magnetic reduction operation, the product would be of a low grade. The relatively small scale of operation would result in a high cost per unit of product.

3.5.3 Other Products

In addition to sulfur compounds, the Carbonization and Solvent Extraction processes produce char, light oil, coal extract, carbon dioxide, and fuel gas. Stanford Research Institute developed estimates of typical selling prices for these products based on geographical locations in the East and Midwest. Prices are listed in the Preliminary Economics section.

The value of char is in doubt. This product would contain all the ash in the refuse feed to the Carbonization process. Because of the original high ash content of the refuse, the char product is not marketable.

The value of the coal extract in the Solvent Extraction process depends upon the specific coal source. It may be used as a chemical, and would then command a higher price than if sold for its fuel value alone.

Section 4

PROCESSING REFUSE FROM COAL PREPARATION

Methods discussed in this section use the entire refuse stream from a coal preparation plant without intermediate treatment to increase the pyrite fraction. The following processes are described:

- Combustion - SO_2 Removal
- Carbonization - Desulfurization
- Solvent Extraction
- Preferential Oxidation

None of these processes has reached commercial status; most are still in development. Originally conceived for run-of-mine coal, these processes encounter special problems when feeding refuse with its very high ash content.

4.1 COMBUSTION - SULFUR OXIDES REMOVAL

Coal refuse can be burned to produce steam for an electric power generator and its sulfur content (sulfur oxides) recovered as sulfuric acid. More than a dozen processes are under active development in the industry for removing sulfur dioxide from power-plant flue gas; none has yet reached commercial status. The major obstacle is economic, not technological. The cost of recovering the SO_2 usually exceeds the value of the sulfur product.

Recovery of SO_2 following combustion of refuse, however, could cost less than recovery from typical power-plant stack gases. Burning coal (with typically a 2.0-percent sulfur content) may produce a stack gas of 0.2 percent SO_2, whereas burning coal refuse (with a 6 to 20 percent sulfur content) may produce a gas of 0.6 to 2.0 percent SO_2. Revenue from the richer SO_2 in the gas could offset the small increase in recovery cost as well as extra costs due to the high ash content of the refuse.

Combustion-Sulfur Oxides Removal combines electric power generation with removal of sulfur oxides from combustion gases. Coal refuse is burned in a steam generator which has the capability to fire a high-ash, high-sulfur fuel. The sulfur present in the refuse is oxidized to sulfur dioxide and a small quantity of sulfur trioxide. The Monsanto Catalytic Oxidation process, on which this study is based, removes the sulfur as sulfuric acid in this air preheating section of the steam generating plant.

The Monsanto process was selected for the following reasons:

- The basic chemistry and technology are sound, and the results of several independent investigators are consistent. For example, the catalytic first stage of the Kiyoura process (developed in Japan) is similar to the Monsanto process.
- It has been tested in small and large pilot plants.
- Sulfuric acid is marketable.
- The process economics would not be penalized by the higher SO_2 concentrations from burning refuse.

The Monsanto process was developed in 1961 by the Pennsylvania Electric Company (Penelec), Air Preheater Company, Monsanto Company, and Research Cottrell Incorporated in a 0.2 Mw(e) pilot plant. Monsanto continued development independently with construction of a 12 Mw(e) pilot plant at the Portland, Pennsylvania, power plant of Metropolitan Edison Company—operation began in 1967. Both power plants were coal-fired. Monsanto claims all rights and patents to the process.

The foregoing process (combustion-sulfur oxides removal), after preliminary technological and economic evaluation, turned out to be the currently most feasible within the constraints of the study. It is therefore examined in some detail in Section 7, and its economic implications expanded there.

4.2 CARBONIZATION-DESULFURIZATION

This technique is under development for run-of-mine coal. Although its application to coal refuse does not appear promising, the technology is of general interest and representative of gasification practice. Carbonization may be defined as the pyrolytic decomposition of coal in the absence of air. The main products of carbonization are:

- Solid semi-coke or char
- Tar
- Combustible gases
- Aqueous liquor

The distribution of the final products depends on the initial composition of the coal and the operating conditions of the process. If the heating is rapid under conditions which prevent decomposition of the primary products, 68 to 87 percent of the heating value of the coal is left in the char, 10 to 23 percent in the tar, and approximately 7 percent in the fixed gases. During normal carbonization, overall desulfurization of coal is usually limited to 50 percent or less.

The technology of carbonization includes hundreds of processes generally characterized by the design of the retort, the type of coal carbonized, and the nature of the char product. Other processes have been proposed which add various compounds, both solid and gaseous, to increase the desulfurization of the char. To increase the yield of liquid and gaseous by-products, processes involving carbonization and subsequent partial hydrogenation have been proposed. The Office of Coal Research is currently sponsoring several developmental projects to convert coal to liquid and gaseous fuels.

Consideration of all such processes for carbonization and desulfurization was beyond the scope of this study. Therefore, after a brief review of the more promising ones, desulfurization and partial gasification was selected for examination. This process appeared to offer the greatest potential for combining a high degree of desulfurization with economic operation.

The specific process is the Consolidation Coal Company CO_2 acceptor process for the devolatization-desulfurization of coal. All associated problems have not been solved; but among the carbonization processes with reasonable potential this one appears to be most advanced. Applied to coal, this process yields a char product having one-percent sulfur content or less, with both elemental sulfur and high-Btu pipeline gas as by-products. The major process steps are:

- Devolatization: The coal is devolatized by contact with hot recycle char.
- Desulfurization: The devolatized mixture is desulfurized by calcined dolomite in the presence of hydrogen.
- Regeneration: The reacted dolomite is removed and regenerated for recycle.

4.2.1 Consol Process

Figure 4-1 shows the basic flow diagram for the Consolidation Coal Company CO_2 acceptor process. Details of the individual process steps follow:

- Coal Preparation: The coal is crushed and carried to a fluid bed dryer using hot (1100 F) gas from the combustion leg. There is no thermal degradation of the coal since the drying is performed at a relatively low temperature (350 F). The fines are collected in a cyclone and blended with the bottoms. The gases are vented through a stack.
- Devolatization: The dry coal is fed into a screw mixer where it is contacted with hot recycled char (1725 F) in the ratio 9:1 (char to coal). The coal fraction is devolatized instantly, and by the time it discharges into the desulfurizer it is nonagglomerating. All of the tar cracks at 1600 to 1700 F to gas containing about 50 percent hydrogen, 20 percent CO, and 20 percent CH_4.
- Desulfurization: The devolatized char is transferred to a fluid-bed desulfurizing vessel. The char is fluidized in a recycled gas containing approximately 50 percent hydrogen. Both the organic and inorganic sulfur in the char react with the hydrogen forming H_2S. These reactions may be represented by the following equations:

  $$R \cdot S + H_2 \rightarrow H_2S + R \qquad \text{(Organic Sulfur)}$$

  $$FeS_2 + 2\,H_2 \rightarrow 2\,H_2S + Fe \qquad \text{(Pyritic Sulfur)}$$

  The equilibrium for these reactions shifts only slightly in the direction of H_2S formation. To obtain a high degree of desulfurization, it is necessary to remove the H_2S as it is formed in order to allow the reactions to go to completion. Calcined dolomite acceptor circulates through the fluid char bed, countercurrent to the gas, and reacts with the H_2S forming a sulfide:

  $$MgO \cdot CaO + H_2S \rightarrow MgO \cdot CaS + H_2$$

By this reaction, the H_2S partial pressure is kept low and a high degree of desulfurization is obtained. After reacting, the acceptor goes to the regenerator.

The gases are cleaned and split into two streams. One stream is converted to high-Btu pipeline gas and the other used as the fluidizing gas.

The char is withdrawn from the reactor continuously; the majority is recycled back into the devolatizer. The remaining char emerges as the product.

- **Regeneration**: Approximately 25 percent of the spent dolomite acceptor is regenerated for recycle in a fluid-bed reactor. Some sulfurized char is burned in air to supply the heat to calcine the dolomite and remove the sulfur as SO_2. The regeneration reaction is:

$$MgO \cdot CaS + 3/2\, O_2 \rightarrow MgO \cdot CaO + SO_2$$

  The gas passes through a cyclone, where the ash and dolomite fines are removed, thence to the sulfur recovery system.

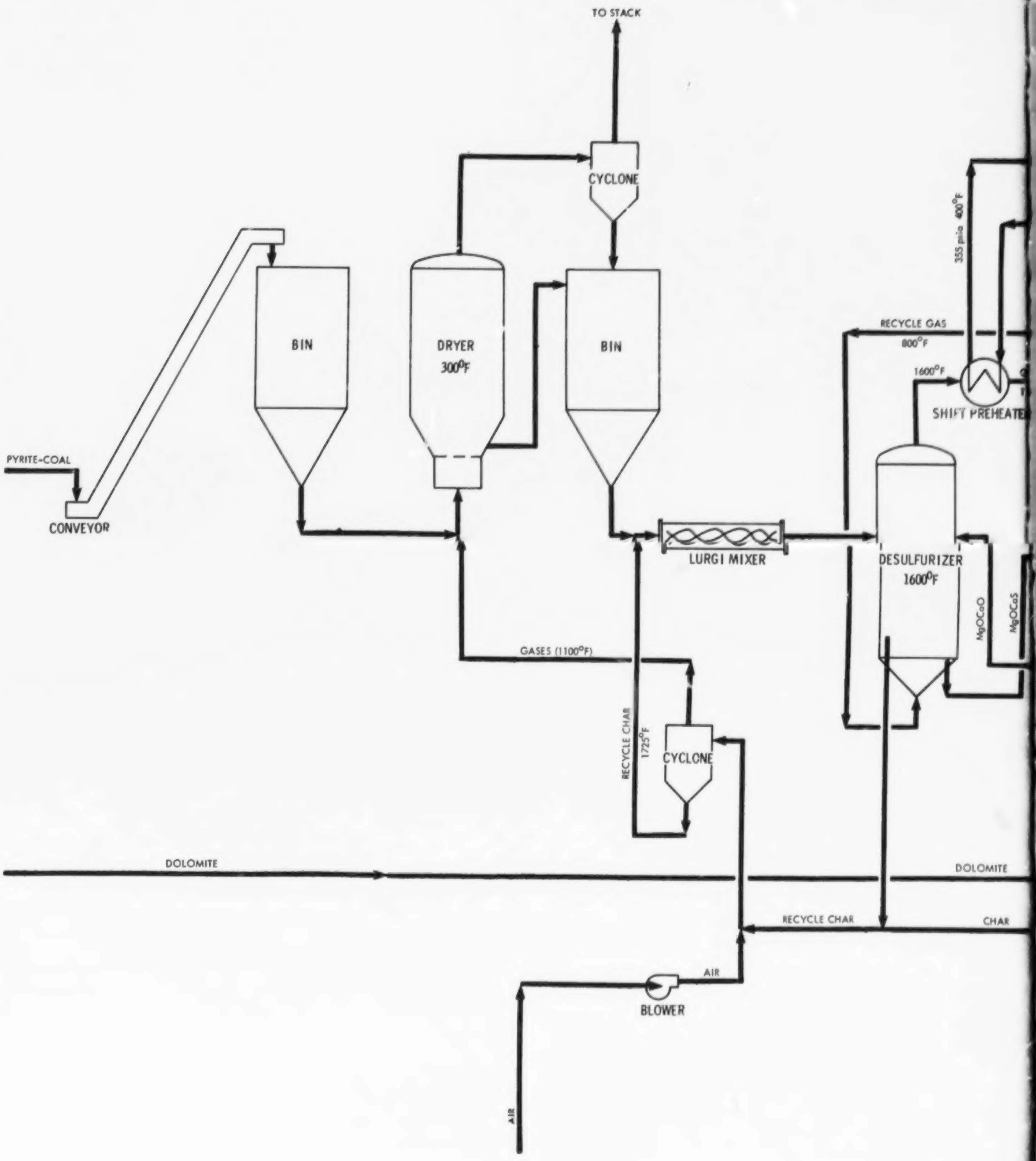
- Sulfur Recovery: The gas passes through a fluid bed of hot char (1500 F) where CO_2 is converted to CO, and is cooled and blended with the CO_2-H_2S from the gas purification section. It is then passed over a bauxite bed where the H_2S and SO_2 react to form elemental sulfur. The reactions can be represented as follows:

$$2\, H_2S + SO_2 \rightarrow S_2 + 2\, H_2O$$

$$SO_2 + 2CO \rightarrow 1/2\, S_2 + 2CO_2$$

  The gases are cooled, and the sulfur collected.

- Gas Purification: The pressurized process gas is heated to 900 F and passed through a shift reactor to obtain the correct ratio (3.2:1) of H_2 to CO for methanation. The gas is then scrubbed with diethanolamine to remove most of the acid gases. The diethanolamine is regenerated and the acid gases (H_2S, CO_2) are sent to the sulfur recovery system. The process gas passes from the scrubber through an iron oxide sponge where H_2S is removed. The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DESULFURIZER
1600°F
MgOCaO
MgOCaS
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DOLOMITE
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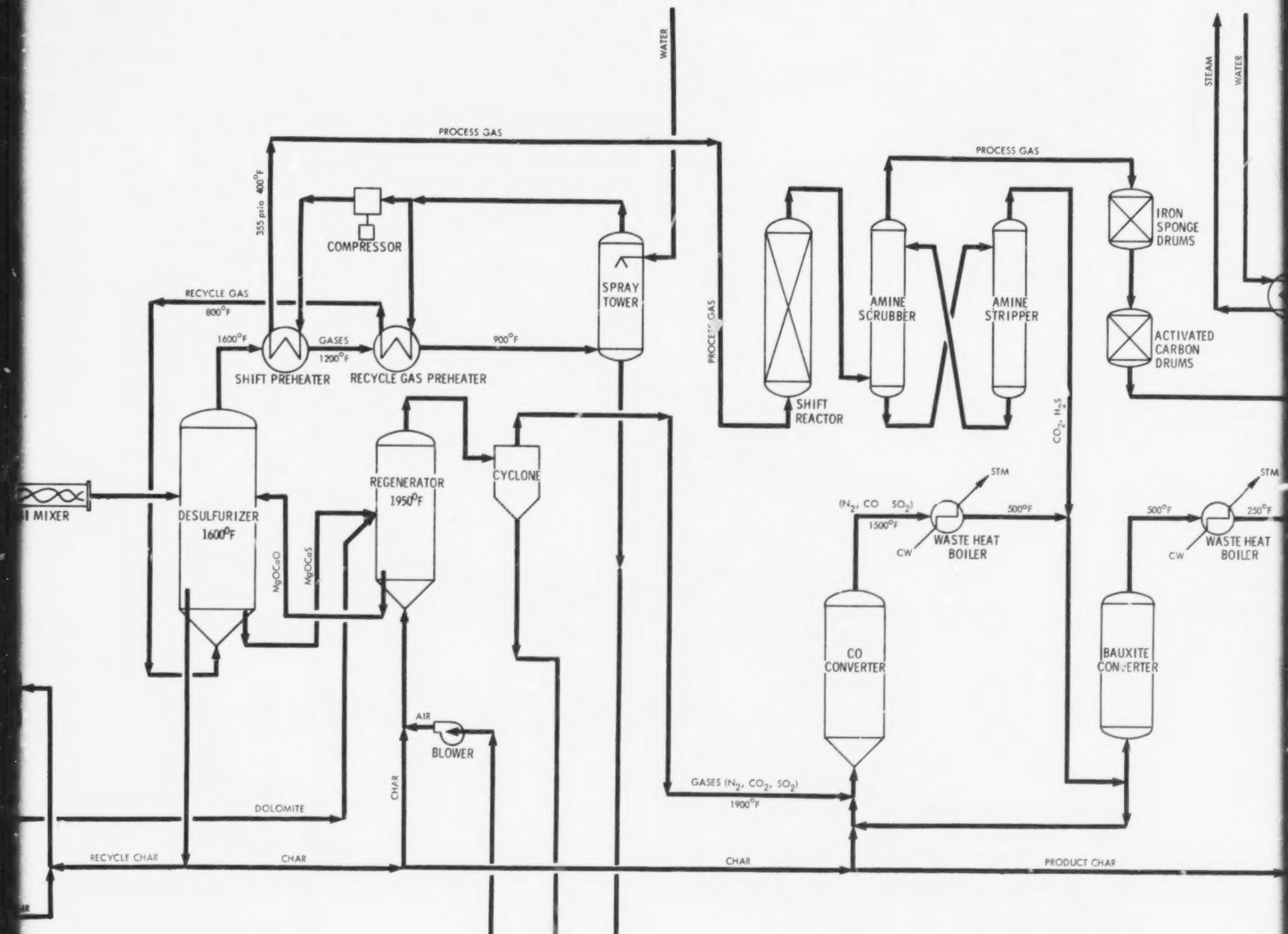
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900°F
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PROCESS GAS
SHIFT REACTOR
PROCESS GAS
AMINE SCRUBBER
AMINE STRIPPER
IRON SPONGE DRUMS
ACTIVATED CARBON DRUMS
STEAM
WATER
MIXER
DESULFURIZER 1600°F
MgOCaO
MgOCaS
REGENERATOR 1950°F
CYCLONE
CO₂, H₂S
(N₂, CO SO₂)
1500°F
STM
WASTE HEAT BOILER
CW
500°F
500°F
STM
250°F
WASTE HEAT BOILER
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BAUXITE CONVERTER
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CHAR
GASES (N₂, CO₂, SO₂)
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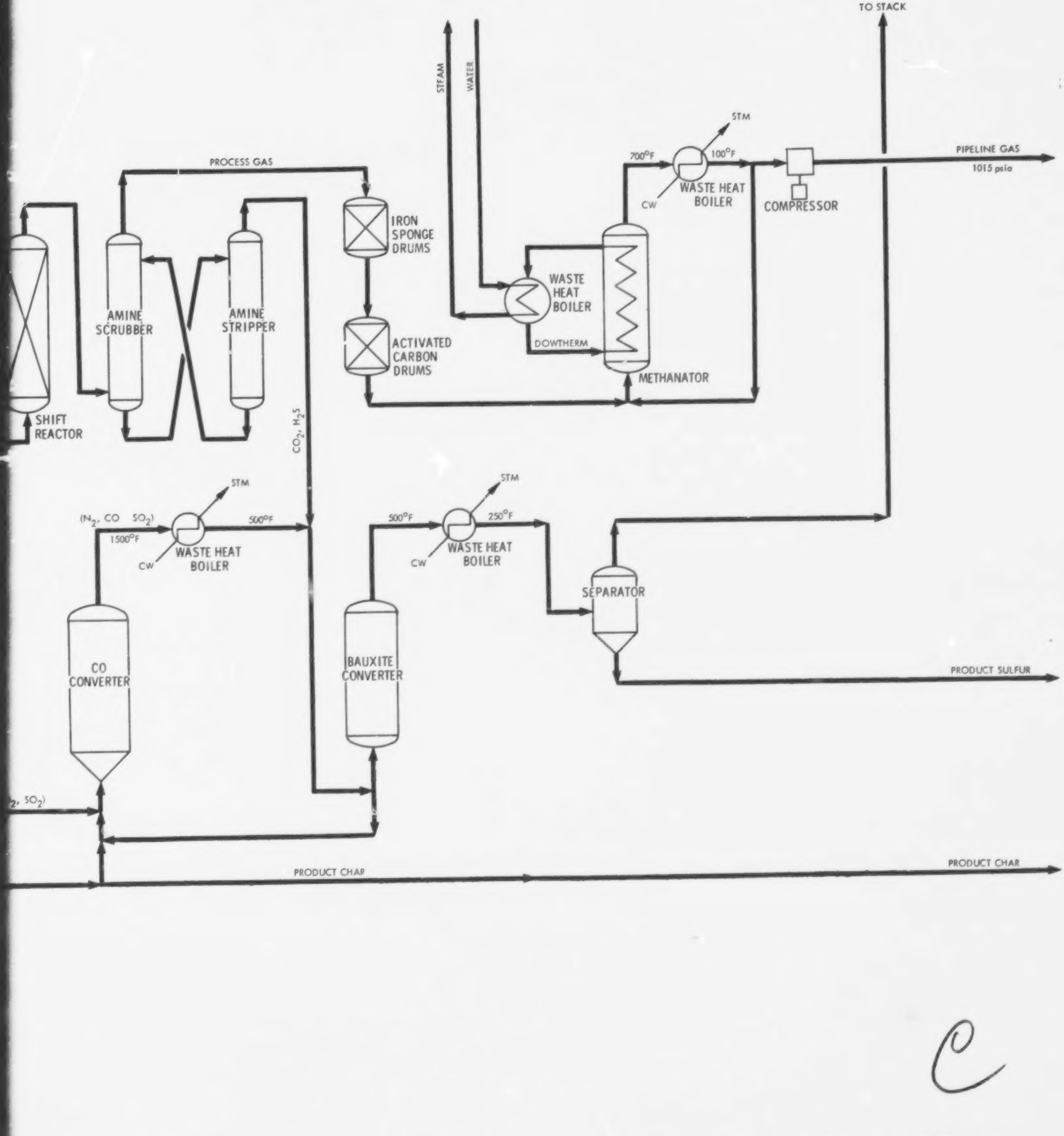
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WASTE HEAT BOILER
AMINE SCRUBBER
AMINE STRIPPER
ACTIVATED CARBON DRUMS
DOWTHERM
METHANATOR
SHIFT REACTOR
CO_2, H_2S
STM
STM
(N_2, CO SO_2)
1500°F
500°F
500°F
250°F
CW
WASTE HEAT BOILER
CW
WASTE HEAT BOILER
SEPARATOR
CO CONVERTER
BAUXITE CONVERTER
PRODUCT SULFUR
(N_2, SO_2)
PRODUCT CHAR
PRODUCT CHAR

C

Figure 4-1. Consol Process Carbonization Desulfurization Process

iron sponge is regenerated by air oxidation. The two reactions can be represented by the following equations:

$$Fe_2O_3 + 3\,H_2S \rightarrow Fe_2S_3 + 3\,H_2O \quad \text{(Removal)}$$

$$2\,Fe_2S_3 + 3\,O_2 \rightarrow 2\,Fe_2O_3 + 6\,S \quad \text{(Regeneration)}$$

The process gas passes through activated carbon beds to remove the last traces of H_2S. It then passes into the methanator where the CO and H_2 react to form methane. Dowtherm vaporizes inside the Raney nickel tube reactor to remove heat of reaction. The gas is finally cooled and compressed.

4.2.2 Adaptation to Refuse Feed

In general, the process appears to be technically feasible, but development work has been confined to processing run-of-mine coal and not coal refuse. Before applying this process to refuse, several problem areas would have to be investigated.

The char product would contain all the ash in the feed material and, because of the high ash content of refuse, it might be unmarketable. The evaluation of char for boiler fuel or other uses would be essential to the economics of the process. In addition, it is possible that the ash would act as a sulfur acceptor and compete with the dolomite for the H_2S. This would increase the sulfur content of the char and make it even more undesirable as a product.

The degree of devolatization of the coal in refuse is particularly critical because the by-product gas is produced exclusively in this step. This phase of the process should be evaluated for various refuse compositions under varying pressures. Higher pressures might increase the devolatization of the coal.

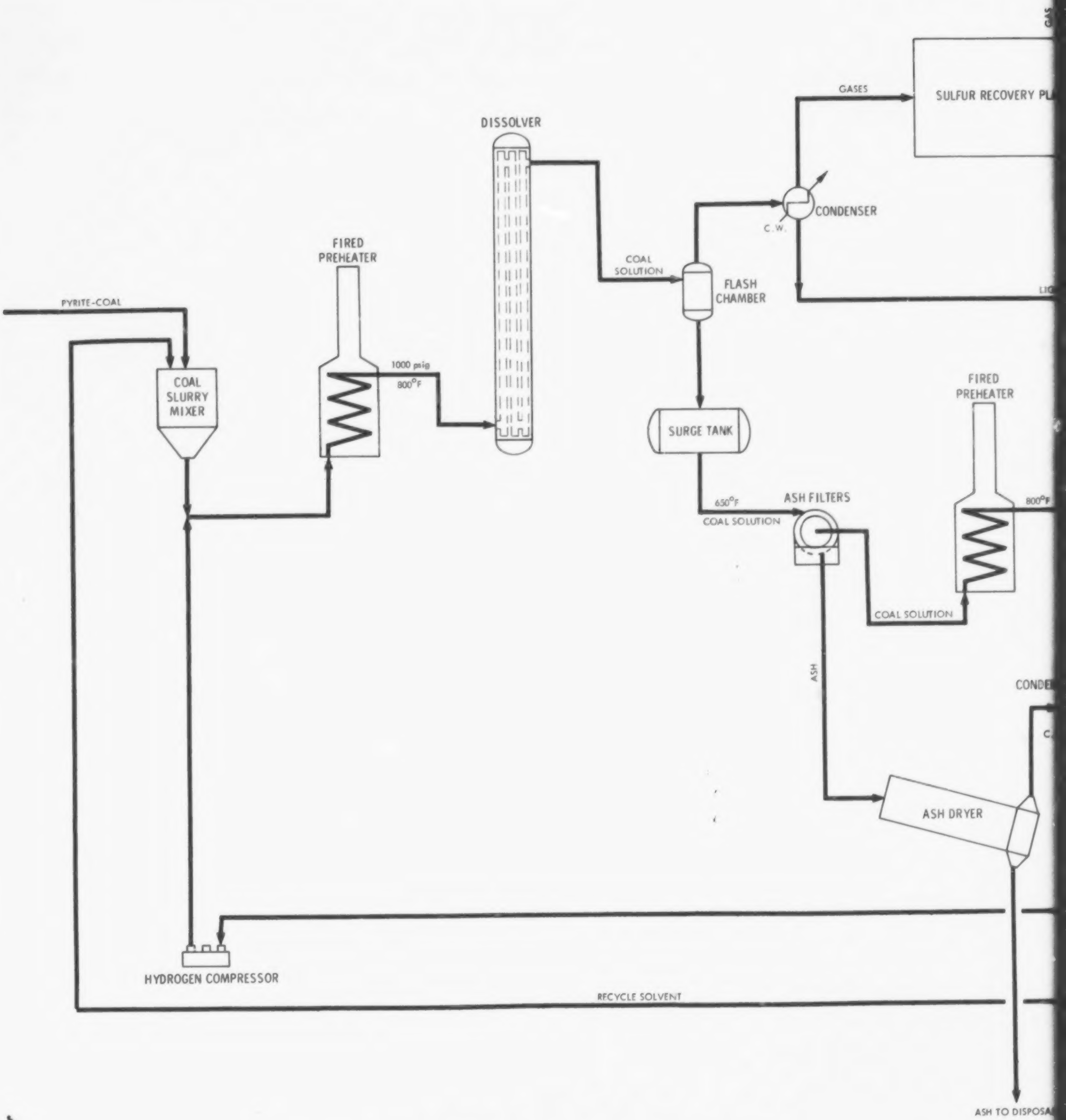
Further investigation of the desulfurization reactions is needed. The possibility of recovering a concentrated H_2S stream from the regenerator due to the reaction ($MgO \cdot CaS + CO_2 + H_2O \rightarrow MgO \cdot CaCO_3 + H_2S$) is important. It is possible that the H_2S recovered from this reaction could be sent directly to a Claus reactor for elemental sulfur recovery. The conditions needed for the recovery of elemental sulfur from the CO-SO_2-H_2S reactions shown in Figure 4-1 should also be investigated.

4.3 SOLVENT EXTRACTION

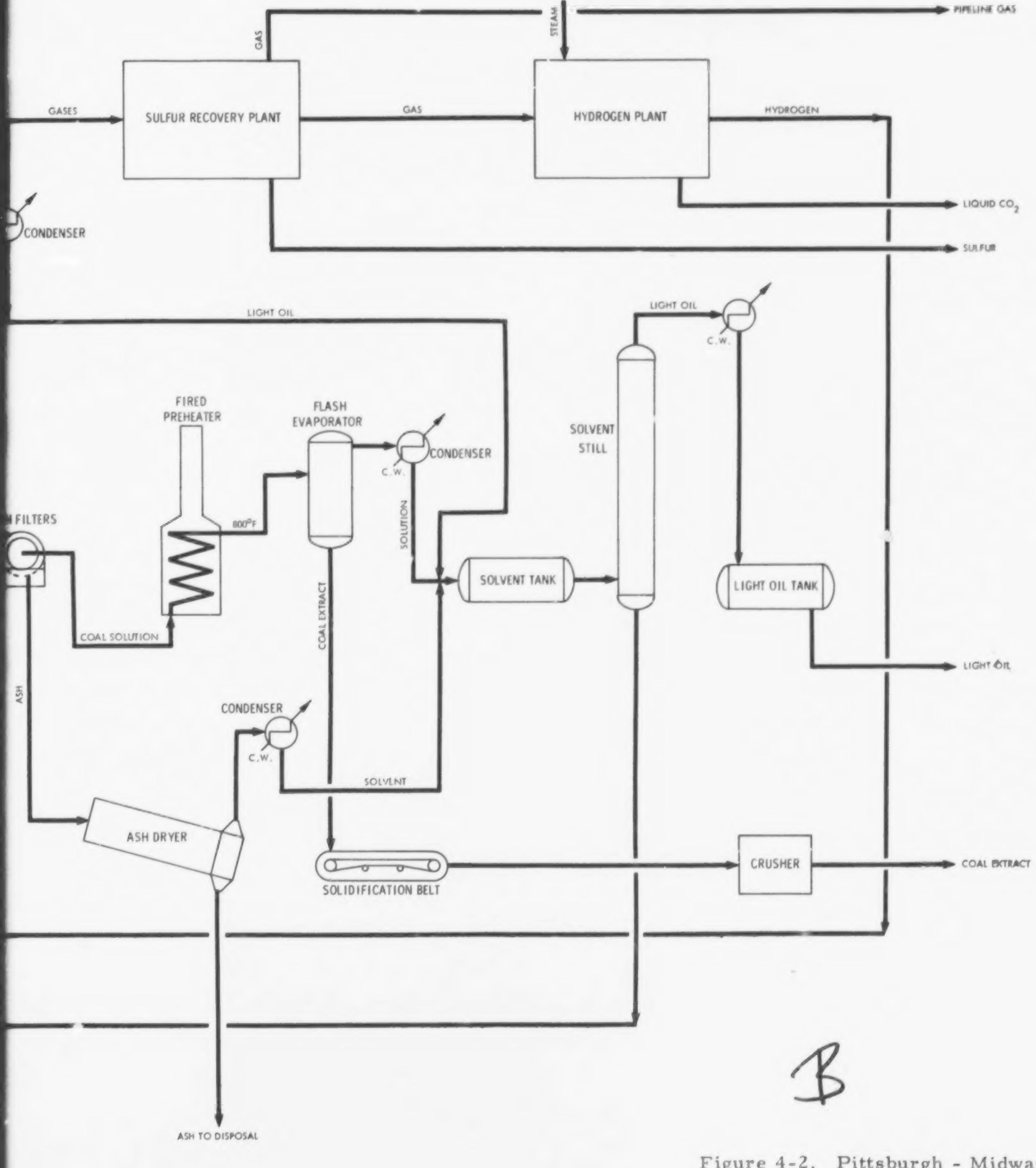
The solvent extraction process uses a solvent, derived from the coal itself, to dissolve the coal in the presence of hydrogen. A solid coal extract which is low in ash and sulfur and has a high heating value is recovered from the solvent.

Processes based upon the solvent extraction of organic matter present in coal to give a low-ash coal extract were developed several years ago. During World War II, the Pott-Broche process was used in Germany. The Japanese have also developed processes which employ the same general principles as the Pott-Broche process to give a low-ash coal extract. Without hydrogenation, the Pott-Broche process required considerable makeup solvent. Operating problems due to solids formation were encountered when hydrogen was added. Difficulties were also encountered in the filtration of the coal solution.

In the United States, both Consolidation Coal Company and the Pittsburgh-Midway Division of Gulf Oil Corporation are currently developing coal solvent extraction processes under the sponsorship of the Office of Coal Research. The Consolidation Coal process, part of Project Gasoline, is currently being evaluated at the Cresap, West Virginia pilot plant. Project Gasoline uses a solvent to dissolve the coal into a liquid fraction


GAS
GASES
SULFUR RECOVERY PL
DISSOLVER
CONDENSER
C.W.
FIRED
PREHEATER
COAL
SOLUTION
FLASH
CHAMBER
LIG
PYRITE-COAL
COAL
SLURRY
MIXER
1000 psig
800°F
SURGE TANK
FIRED
PREHEATER
650°F
ASH FILTERS
COAL SOLUTION
800°F
COAL SOLUTION
ASH
CONDE
C
ASH DRYER
HYDROGEN COMPRESSOR
RECYCLE SOLVENT
ASH TO DISPOSA

A


PIPELINE GAS
GAS
STEAM
GASES
SULFUR RECOVERY PLANT
GAS
HYDROGEN PLANT
HYDROGEN
LIQUID CO_2
CONDENSER
SULFUR
LIGHT OIL
LIGHT OIL
C.W.
FIRED PREHEATER
FLASH EVAPORATOR
SOLVENT STILL
C.W.
CONDENSER
FILTERS
800°F
SOLUTION
SOLVENT TANK
LIGHT OIL TANK
COAL EXTRACT
COAL SOLUTION
LIGHT OIL
ASH
CONDENSER
C.W.
SOLVENT
ASH DRYER
SOLIDIFICATION BELT
CRUSHER
COAL EXTRACT
ASH TO DISPOSAL

B

Figure 4-2. Pittsburgh - Midway Solvent Extraction Process

which can be transformed into gasoline. Pittsburgh-Midway has demonstrated the technical feasibility of processing coal by solvent extraction to produce a low-ash, low-sulfur product. The company is presently developing a pilot plant to thoroughly investigate the operating parameters which affect the process.

It was decided to evaluate the Pittsburgh-Midway process for solvent extraction of refuse because of the potential for recovery of high-value products. The coal extract recovered could be utilized as a chemical and command a higher price than if it were sold for its fuel value alone. The major process steps are:

- Solution of Coal: The coal is dissolved in solvent derived from the coal itself in the presence of hydrogen.
- Filtration: The coal solution is filtered to remove the undissolved mineral matter.
- Solvent Recovery: The solvent is removed for recycle by flash evaporation, and the de-ashed coal extract recovered.

4.3.1 Solvent Extraction Process

Figure 4-2 shows the basic process flow diagram for the Pittsburgh-Midway process. Details of the individual process steps follow:

- Coal Solution: The pulverized coal is slurried with recycled coal solvent to give approximately a 2:1 solvent-to-coal ratio. The coal slurry is then mixed with hydrogen at approximately 1000 psig, heated to 800 F in a fired preheater, and fed to the dissolver. The dissolver is an isothermal, unagitated, tubular reactor which may operate adiabatically; however, indications are that it will have to be heated to maintain isothermal conditions.

  The solution of the coal in the dissolver produces a broad range of products. These include a considerable quantity of material which is identical to the original solvent. In addition, products of low molecular weight such as methane, ethane, and propane are found, as well as numerous aromatic and aliphatic hydrocarbons in the light oil fraction.

Both the labile pyritic and organic sulfur present in the coal slurry react with the hydrogen in the dissolver to give H_2S. The reactions can be represented as follows:

Organic Sulfur: $R \cdot S + H_2 \rightarrow H_2S + R \cdot$ (I)

Pyrite Sulfur: $FeS_2 + H_2 \rightarrow FeS + H_2S$ (II)

Approximately 75 percent of the organic sulfur is removed by Reaction I. About one-half of the pyrite-sulfur is converted to H_2S by Reaction II, and the remaining sulfur is left as insoluble FeS in the coal solution. The H_2S in the process gas stream is separated along with the solvent vapor ahead of the filters in a flash chamber. The process gas then goes to a gas cleaning and sulfur recovery system. The solvent vapor is condensed and sent to a solvent separation system.

- Filtration: The coal solution is filtered at approximately 650 F to remove the undissolved coal and mineral matter. The ash is dried to remove the entrained solvent and is then removed for disposal or subsequent use, depending on the value.
- Extract Recovery: The filtrate, which has been heated to 800 F in the preheater, is flash evaporated to recover the solvent. The liquid coal extract is solidified and crushed.
- Solvent Distillation: The solvent is distilled to recover the light oil fraction; the heavier fraction is returned to the slurry mixer.
- Sulfur Recovery: The acid gases are removed from the process gas stream, and the sulfur is recovered in a Claus plant. A portion of the cleaned gas stream is used for hydrogen production and the remainder is by-product gas.
- Hydrogen Production: Process gas is steam-reformed to produce the hydrogen needed in the process. Liquid CO_2 is recovered as a by-product.

4.3.2 Adaptation to Refuse Feed

Because of the developmental status of the process, it is difficult at the present time to evaluate accurately its use for treating refuse. It would appear that solvent processing can be applied to coal refuse; however, a complete economic evaluation cannot be made without additional information. Further investigations are needed with respect to the efficiency of this process. These relate to the heat and material balances, solvent recovery from filter cake, and the filtration system.

The differences between ROM coal and coal refuse are significant, and should be evaluated with respect to the operation of the process. Some of the parameters which should be evaluated are listed below:

- Sulfur: A knowledge of the mechanism and degree of liberation of the pyritic and organic sulfur fractions is needed, as well as the ultimate distribution of sulfur in the various phases of the process. The effect of increased pyrite feed on the sulfur balance should also be evaluated. Important factors are the possible breakdown of pyrite into unfilterable particles, hydrogen partial pressure, H_2S partial pressure, temperature, and retention time.
- Hydrogen: The determination of the hydrogen consumption for coal depolymerization, sulfur liberation, and process gas formation is essential. The effect of hydrogen partial pressure, pyrite content of feed, temperature, and retention time should be evaluated.
- Heat Requirement: A definitive study of the heat requirements for the process is essential to the development of a full-scale process. Heat required for solution of coal is a key parameter.
- Product Yield: Evaluation of coal solution product yield with variations in feed material composition or operating conditions is needed. The coal in the refuse mixture may be insoluble under normal operating conditions. Also, the sensitivity of product yield to changes in operating conditions may affect the economic evaluation

> of the process. Temperature, retention time, pressure, and the mineral matter and moisture content of the feed are important factors.

Coal-derived solvent is removed from the filter cake by displacement. This is effected by using a solvent of lighter boiling point. The degree of removal, as well as the effect of increased mineral matter on solvent removal should be investigated under continuous operating conditions. The recovery and recycle of the solvent is an essential part of the process.

The filtration efficiency is important if the coal extract is to be used for applications other than fuel for power generation. The effect of increased mineral matter and pyrite on filtration should be evaluated under conditions of varying ratios of pyrite to mineral matter. In addition, the use of centrifuges should be investigated, although the increased ash and sulfur content of the coal extract might limit this use.

4.4 PREFERENTIAL OXIDATION

If a high-sulfur coal reacts with a limited amount of air, at least part of the sulfur should be removed by oxidation. Such a process, however, should react a maximum of sulfur without burning excessive amounts of coal. No commercial process currently exists that does this, and very little experimental work has been published.

Bench-scale experiments have shown that it is possible to oxidize the pyritic sulfur in coal preferentially without burning an excessive amount of the coal. In one set of experiments, a 4.9-percent sulfur coal was fluidized at 800 F by a mixture of air, steam, and nitrogen. About half the sulfur was present in pyritic form and half in organic. The treatment removed approximately 80 percent of the pyritic sulfur, but only about 10 percent of the organic sulfur.

The experiments showed that the oxygen content of the fluidizing gas was the key factor to the preferential oxidation of sulfur over carbon. Other results showed that above 800 F there was excessive gasification and loss of char yield, and that a residence time of approximately 18 minutes was required. The 800 F temperature condition is considerably below the 1750 F temperature for commercial pyrite roasting, which means the reaction rate is lower. Free-energy calculations show that the reaction is thermodynamically feasible at 800 F.

The major problem in applying a single-stage, preferential oxidation to pyrite-coal concerns control of heat of reaction and temperature. Pyrite oxidizes so rapidly that it would be an extremely difficult engineering problem to maintain the temperature around 800 F.

Section 5

PROCESSING PYRITE CONCENTRATES

Methods discussed in this section were designed for commercial grades of pyrite ranging from 40 to 50 percent sulfur. Since the refuse stream may contain from 6 to 20 percent sulfur, it must be concentrated to make it an appropriate pyrite feed.

Roasting of pyrite concentrates is a fully developed technology which has existed for many years. Iron pyrites are mined principally for their sulfur value, which is usually recovered by roasting the ore to sulfur dioxide and iron oxide. The sulfur dioxide is converted to sulfuric acid. Minor amounts of base metals are often associated with pyrite (e. g., the minerals chalcopyrite ($Cu_2S \cdot FeS \cdot FeS_2$), blende (ZnS), and galena (PbS)). Iron oxide, nonferrous, and precious metals may or may not be recovered, depending on economics. A few commercial processes recover elemental sulfur.

Table 5-1 describes the various furnaces used for roasting pyrite. Fluid beds are perhaps the strongest competitors for new installation for low-arsenic pyrite. Multiple hearths are expensive but capable of handling high-arsenic pyrite, while rotary kilns and flash furnaces are less flexible and require special feeds. Each ore body has its own specific requirements which determine the roaster design.

Most pyrite roasting operations produce only SO_2 for conversion to acid. For example, in 1965 approximately ten million tons of sulfur equivalent were produced from pyrite. Only about 161,000 tons of elemental sulfur were recovered. However, this small production of elemental sulfur from pyrite does not mean that such processes are unimportant. Considerable research and development effort is directed toward processes to recover elemental sulfur.

Three pyrite processes were picked for more detailed description. All three have reached commercial status, and it may be possible to adapt them to pyrite-coal mixtures.

5.1 FLUID-BED ROASTING

Fluid-bed pyrite roasting and acid recovery is a widely used commercial-scale technique. The first installation for roasting a sulfide ore in a fluid bed was in Ontario, Canada, at the Cochenour Williams Gold Mines, Ltd. This plant, installed in 1946, eliminated sulfur and arsenic from the ore to free the gold for recovery by cyanidation.

The Dorr-Oliver Company claims to have sold 104 fluid-bed systems for roasting pyrites. The process is used by the Tennessee Copper Company at Copper Hill, Tennessee; by the Bethlehem Steel Company at Sparrows Point, Maryland; and by the Kennecott Copper Company at Hayden, Arizona. Esso Standard Fertilizer and Chemical Company roasts about 600 tons per day of pyrite concentrates by fluid beds in the Philippines.

The Montecatini plant in Italy, which came onstream in 1964, uses the most advanced fluid-bed technology for roasting pyrites. The plant was designed to feed 1100 metric tons per day of pyrite containing 35 to 39 percent sulfur and 35 to 39 percent iron. It produces 1100 metric tons per day of sulfuric acid (100 percent basis), 550 metric tons per day of

Table

FURNACES FOR RO

	Design and Operations	
FLUID BED	The furnace consists of a cylindrical, brick lined, steel shell which is closed at the bottom by a grate. Air or other fluidizing gas flows up through the grate from a windbox and into the bed of solids. Gas velocity is controlled to keep the bed turbulent. Pyrite (max. 1/4 in.) is fed continuously as a dry solid or as a 70% solid slurry in water. The coarse fraction of the roasted pyrite leaves the furnace through an overflow pipe; the fines carry over with the gases and are recovered.	● Inti acti fer ● A f put ● Pro ● The hea ● Con low les
MULTIPLE HEARTHS	The furnace consists of a cylindrical, brick-lined, steel shell with a series of hearths stacked vertically on the same axis. Each hearth has rabble arms fixed on a rotating vertical shaft. Ground pyrite descends from hearth countercurrent to ascending gases and finally leaves the bottom of the furnace as calcined oxide.	● Thi put, ● It i ● It c furn
SUSPENSION ROASTER	The furnace consists of a brick-lined, steel shell designed as a combustion chamber. Very fine particles of pyrite are injected into the upper part of the combustion chamber. The combustion gases flow at a velocity which will suspend the pyrite particles. Air or oxygen can be used to burn the pyrite to SO_2 and iron oxide. Larger particles of the cinders settle to the hopper bottom of the combustion chamber. Fines are recovered from off-gases.	● Thi furn ● Inti rat ● Off- ● Goo
ROTARY KILN	Similar to conventional cement kilns.	● Thi ● The anc ● Cin ● Util

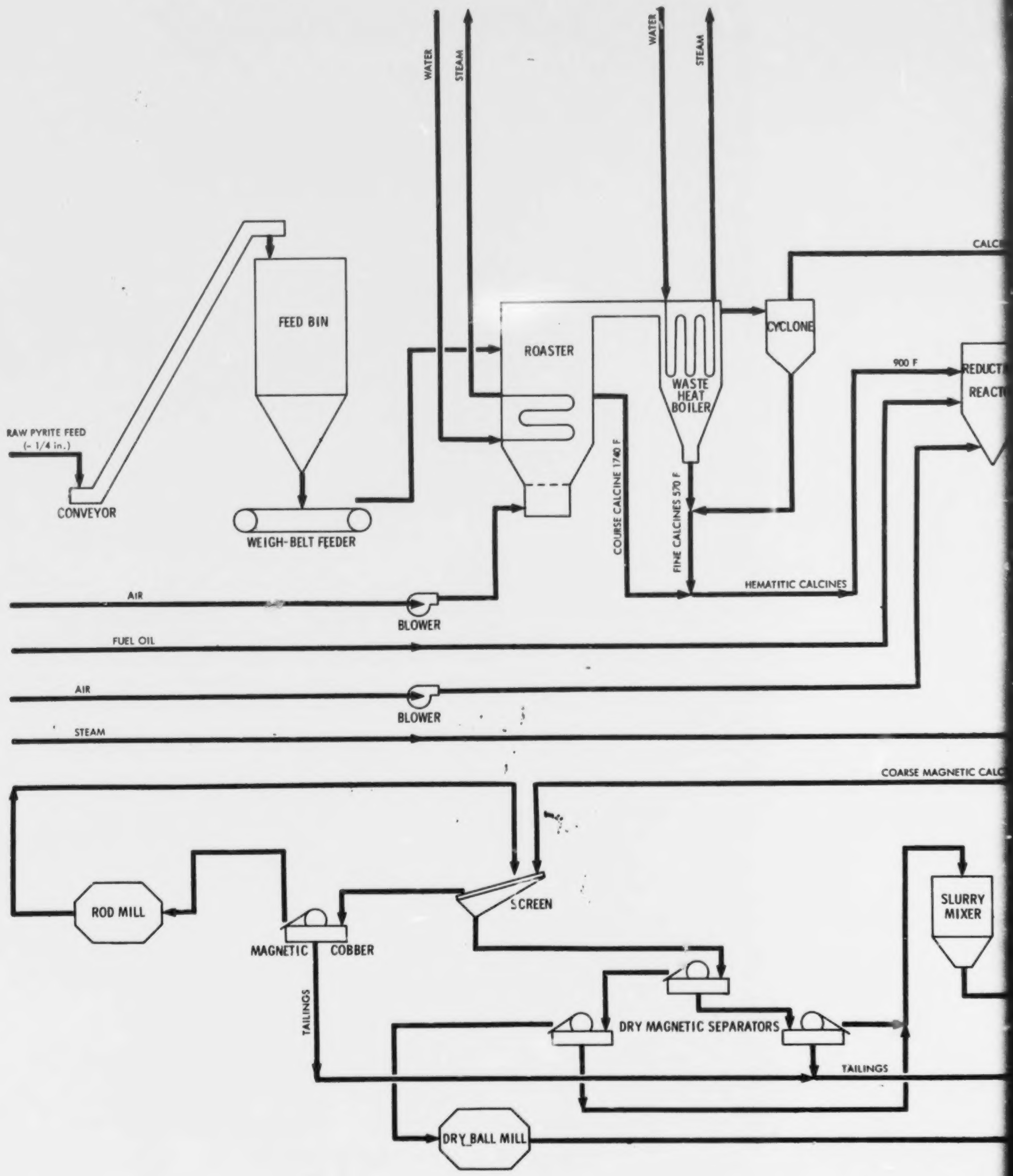
A

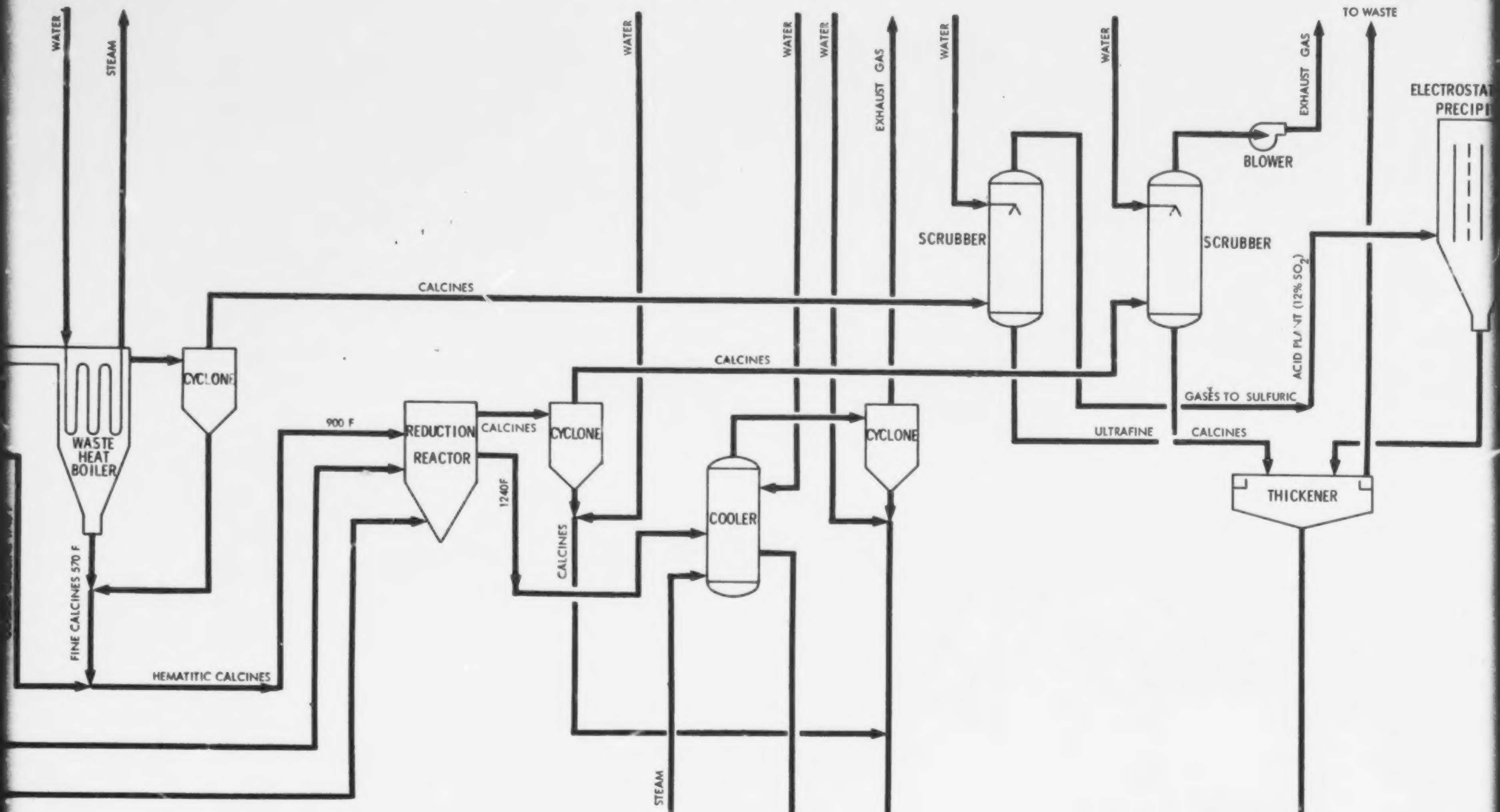
Table 5-1

FOR ROASTING PYRITE

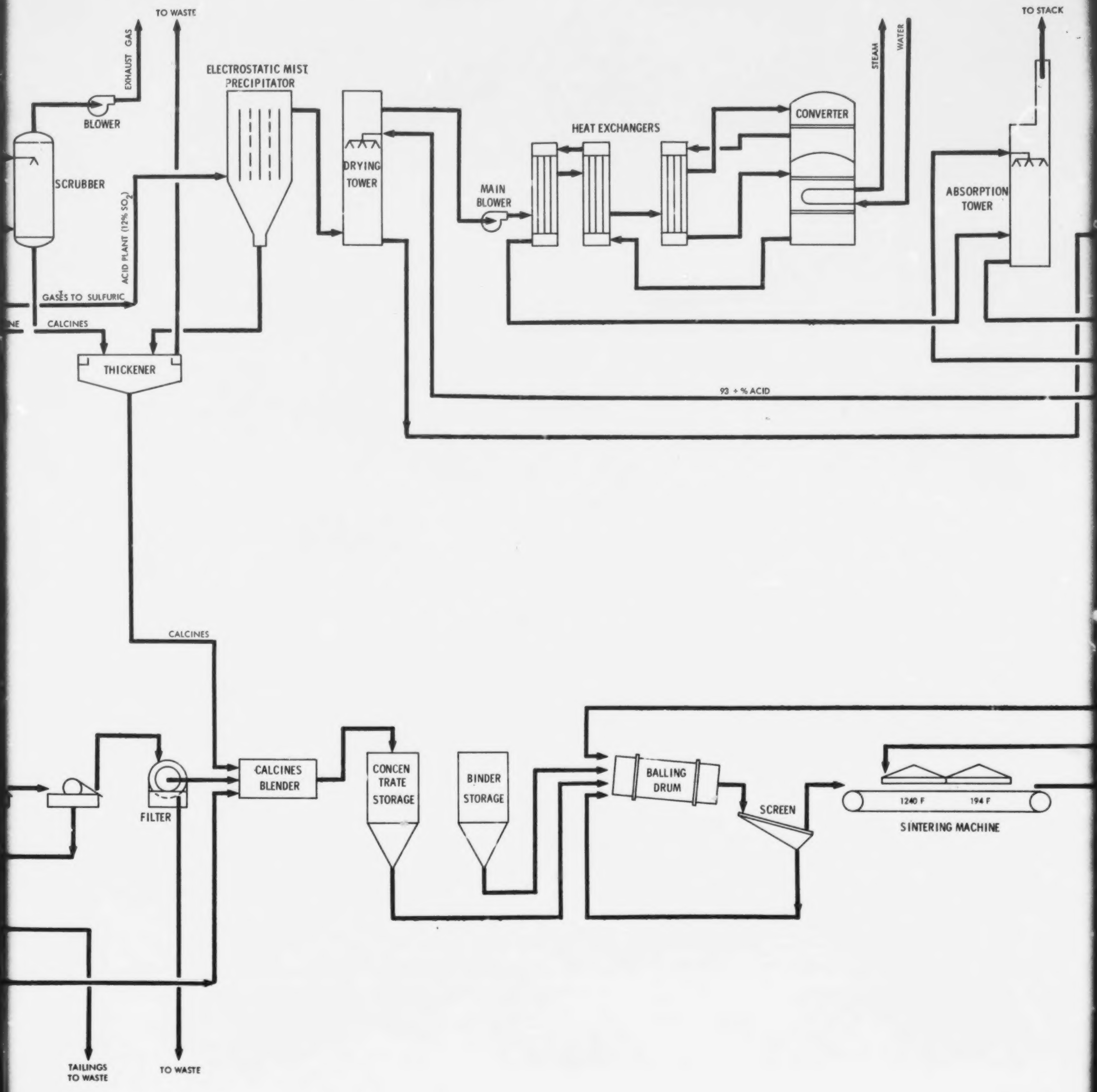
Characteristics	Limitations
Intimate gas-solids contact increases reaction rates and gives excellent heat transfer and bed temperature control. A fluid bed is capable of high rates of throughput per unit of hearth area. Product gas has high SO_2 content. The design permits convenient addition of heat recovery units. Compared to multihearths, fluid beds have low capital and operating costs and require less floor space	● A low-arsenic pyrite feed is required; the cinders will otherwise be too high in arsenic for blast-furnace feed De-velopment work shows that this problem may be solved.
This furnace is capable of high rates of throughput, but at the expense of large hearth area. It is good for removing arsenic from residue. It can produce cinder suitable for a steel blast-furnace but with further processing.	● Product gas has lower SO_2 content (8%) than fluid bed. Recent designers claim 11% SO_2 ● Heat recovery is lower than fluid-bed. New designs claim improvements.
This furnace is capable of high capacity per unit furnace area. Intimate gas-solids contact gives fast reaction rates and good heat transfer. Off-gases have high SO_2 content (14%). Good heat recovery is possible.	● It does not produce low-arsenic cinders from arsenical pyrite (0.5% As). ● It is not suitable for sulfating roast. ● Pulverized feed, such as flotation concentrates, is required.
This furnace is capable of high capacity. The furnace design is simple, with low maintainance. Cinders have low sulfur content. Utility and labor costs are low.	● Capital cost is high. ● Off-gas has low concentration of SO_2 (7-8%). ● There is a high dust carry-over.

B

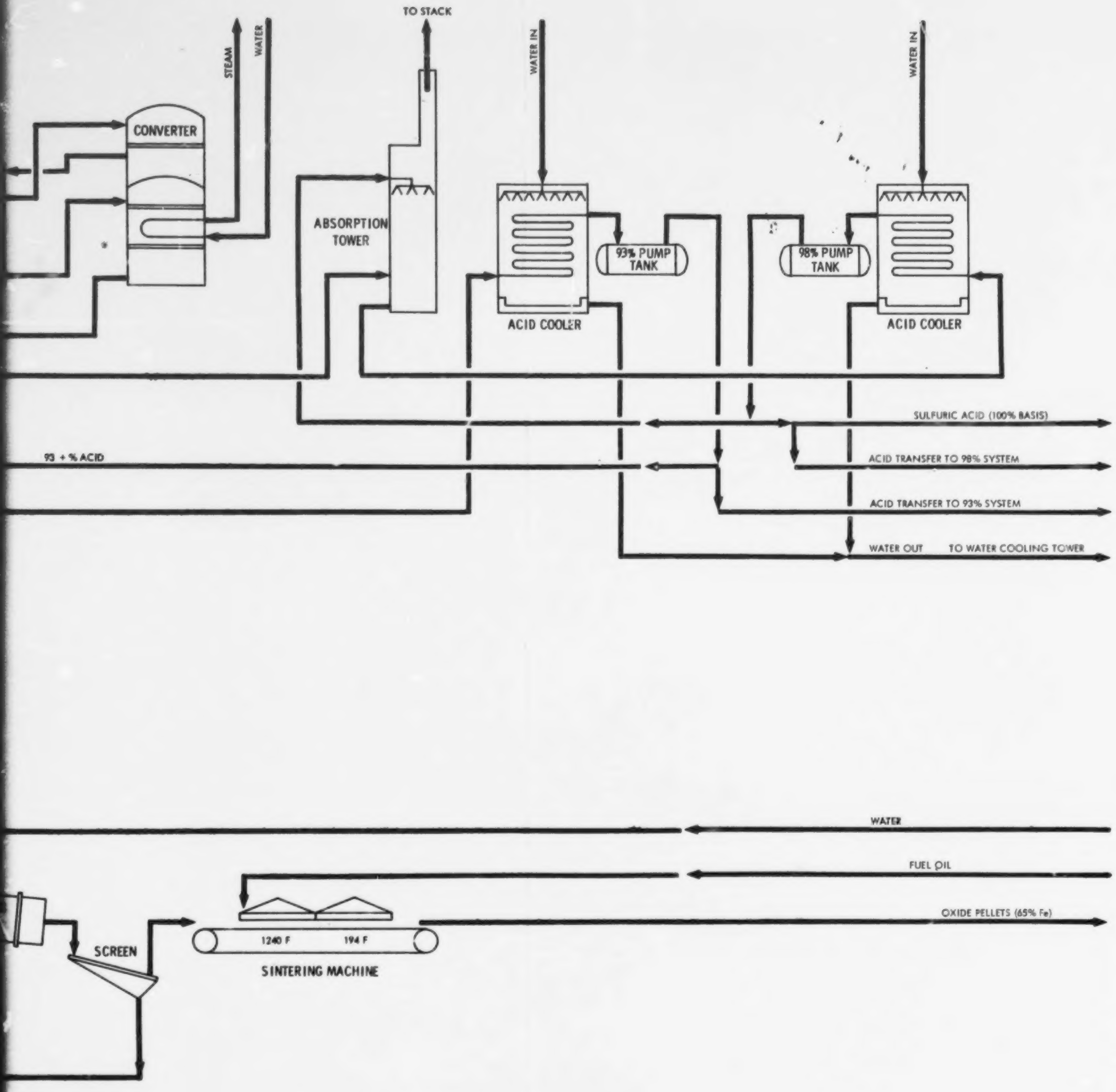

WATER
STEAM
WATER
STEAM
CALCI
FEED BIN
ROASTER
WASTE HEAT BOILER
CYCLONE
900 F
REDUCTI
REACT
RAW PYRITE FEED (- 1/4 in.)
CONVEYOR
WEIGH-BELT FEEDER
COURSE CALCINE 1740 F
FINE CALCINES 570 F
HEMATITIC CALCINES
AIR
BLOWER
FUEL OIL
AIR
BLOWER
STEAM
COARSE MAGNETIC CALC
ROD MILL
SCREEN
MAGNETIC COBBER
TAILINGS
SLURRY MIXER
DRY MAGNETIC SEPARATORS
TAILINGS
DRY BALL MILL


WATER
STEAM
WATER
WATER
WATER
EXHAUST GAS
WATER
WATER
EXHAUST GAS
TO WASTE
BLOWER
ELECTROSTAT
PRECIPI
SCRUBBER
SCRUBBER
ACID PLANT (12% SO_2)
CALCINES
CALCINES
GASES TO SULFURIC
CYCLONE
WASTE HEAT BOILER
900 F
REDUCTION REACTOR
CALCINES
CYCLONE
1240F
CYCLONE
COOLER
ULTRAFINE
CALCINES
THICKENER
CALCINES
FINE CALCINES 570 F
HEMATITIC CALCINES
STEAM
COARSE MAGNETIC CALCINES 356 F
CALCINES
FINE MAGNETIC CALCINES
SLURRY MIXER
WET BALL MILL
WET MAGNETIC SEPARATORS
FILTER
CALC
BLE
RY MAGNETIC SEPARATORS
TAILINGS
TAILINGS TO WASTE
TO WASTE

B


TO WASTE
EXHAUST GAS
BLOWER
SCRUBBER
ELECTROSTATIC MIST PRECIPITATOR
ACID PLANT (12% SO_2)
GASES TO SULFURIC
CALCINES
THICKENER
DRYING TOWER
MAIN BLOWER
HEAT EXCHANGERS
CONVERTER
STEAM
WATER
TO STACK
ABSORPTION TOWER
93 + % ACID
CALCINES
FILTER
CALCINES BLENDER
CONCEN TRATE STORAGE
BINDER STORAGE
BALLING DRUM
SCREEN
1240 F
194 F
SINTERING MACHINE
TAILINGS TO WASTE
TO WASTE

C


TO STACK
STEAM
WATER
WATER IN
WATER IN
CONVERTER
ABSORPTION TOWER
93% PUMP TANK
98% PUMP TANK
ACID COOLER
ACID COOLER
SULFURIC ACID (100% BASIS)
93 + % ACID
ACID TRANSFER TO 98% SYSTEM
ACID TRANSFER TO 93% SYSTEM
WATER OUT
TO WATER COOLING TOWER
WATER
FUEL OIL
OXIDE PELLETS (65% Fe)
1240 F
194 F
SCREEN
SINTERING MACHINE

D

Figure 5-1. Fluid-Bed Roasting Montecatini Process

high purity iron oxide pellets (66 to 67 percent Fe), and 570,000 kilowatt-hours per day of electricity from waste heat. Montecatini is now completing construction that will double the above plant capacity. The major process steps are:

- Pyrite Roast: Pyrite is dead-roasted in a fluid-bed roaster to produce SO_2 and a low-sulfur hematitic calcine.
- Iron-Oxide Reduction: The hematite (Fe_2O_3) in the roaster calcine is reduced to magnetite (Fe_3O_4) by direct injection of fuel oil in a fluid-bed roaster.
- Magnetic Separation: A combination of wet and dry magnetic separators removes the gangue from the magnetic calcines.
- Product Recovery: The iron oxide concentrates from magnetic separation are pelletized to give a high-grade iron ore product (66 to 67 percent Fe). Sulfur dioxide is converted to sulfuric acid in a conventional contact plant. Waste heat from the pyrite roast generates electricity.

5.1.1 Montecatini Process

Figure 5-1 shows the process flow. A description of the major process steps follows.

- Roasting: An electronic weigh-belt feeder delivers run-of-mine pyrite, crushed to 1/4 inch, to the fluid-bed roaster. The pyrite burns autogenously to hematite with 10 to 20 percent excess air at about 1740 F according to the reaction:

  $$2\ FeS_2 + 5\text{-}1/2\ O_2 \rightarrow Fe_2O_3 + 4\ SO_2$$

  Exothermic heat of reaction amounts to about 2000 Btu per pound of pyrite. The off-gas with 12 percent SO_2 discharges through a waste heat boiler to produce steam at 840 F and 590 psig. From the waste heat boiler, the gases are further cleaned by a cyclone and scrubber before going to the acid plant for conversion to 98 percent H_2SO_4.

The hematitic calcines are obtained as a coarse fraction from the fluid bed, as a fine-sized fraction from the cyclones, and as an ultrafine fraction from the scrubber. This ultrafine fraction does not need further upgrading. It is dewatered in a thickener and goes directly to the pelletizing plant.

To minimize heat losses, calcines from the roaster bed and the waste heat boiler go immediately to the adjacent reduction reactor.

- Reduction: Hematite is reduced to magnetite in a fluid-bed reactor at about 1240 F. Fluidization is maintained by a reducing gas produced by partial combustion of fuel oil injected directly into the bed. Both carbon monoxide and hydrogen are present in the gases and react with hematite:

$$3\,Fe_2O_3 + CO \rightarrow 2\,Fe_3O_4 + CO_2$$

$$3\,Fe_2O_3 + H_2 \rightarrow 2\,Fe_3O_4 + H_2O$$

The reduction chamber is designed and operated to produce a 150-mesh products classification, which is optimum for the subsequent dry and wet magnetic separations. The fine fraction, which carries over into the cyclones, amounts to about 25 percent of the feed to the reduction reactor. Coarse magnetic calcine is suitable for dry magnetic separators; fine calcines must be handled in a wet magnetic scheme.

- Magnetic Separation: The coarse fraction of the magnetic calcine passes from the reduction reactor into a fluid-bed cooler (fluidized by steam) and cools to about 350 F. This dust-free fraction of the calcines, after a slight grinding, feeds to the dry magnetic circuit. The gangue discard is slurried and pumped to waste.

The fine fraction, collected from the cyclones, is quenched in water and fed to the wet magnetic circuit. Gangue is pumped to waste and the iron oxide concentrate is dewatered by a rotary filter.

- Pelletizing: The dry and wet concentrates combine to give a final concentrate with about 12 percent moisture. Enough water is added to raise the moisture content to 14 percent before feeding to the balling drum. A straight line grate processes the green pellets through the stages of drying, preheating, firing to 2400 F, and cooling to 190 F. The product iron oxide pellets assay 66 to 67 percent Fe.

The Montecatini process has the advantages inherent in fluid-bed technology. Fluid beds require less capital investment, less maintenance, and less space than multihearth shaft furnaces. Favorable reaction kinetics are obtained due to close gas-solid contact at high mass flow; this results in good heat-transfer rates and close bed temperature control.

The second step in the Montecatini process, the magnetizing roast, permits upgrading the calcine. But even more important and essential, it removes sulfur from a percent or so down to 0.1 percent or less to make the pellets attractive for blast furnace feed. Sulfur removal in the blast furnace is expensive.

The highly developed technology of the Montecatini/Dorr-Oliver process makes it unique, in that a high-grade iron ore is recovered. The flowsheet shows that considerable equipment investment is required to upgrade the calcine into a marketable product. The biggest disadvantage of the process, however, is that it does not produce elemental sulfur, which limits the marketing area.

Most commercial pyrite roasting processes do not recover a high-grade iron by-product but may sell the cinder to makers of building block. This would be especially true for pyrite-coal due to the ash content. Figure 5-2 shows a simpler processing scheme typical of fluid-bed roasting without iron recovery. It consists of a feed delivery system, roasters with waste heat boilers, gas cleaning system, and waste calcine disposal. Although the illustration shows a dry feed system, slurry feed can be used and probably is used more often than the dry.

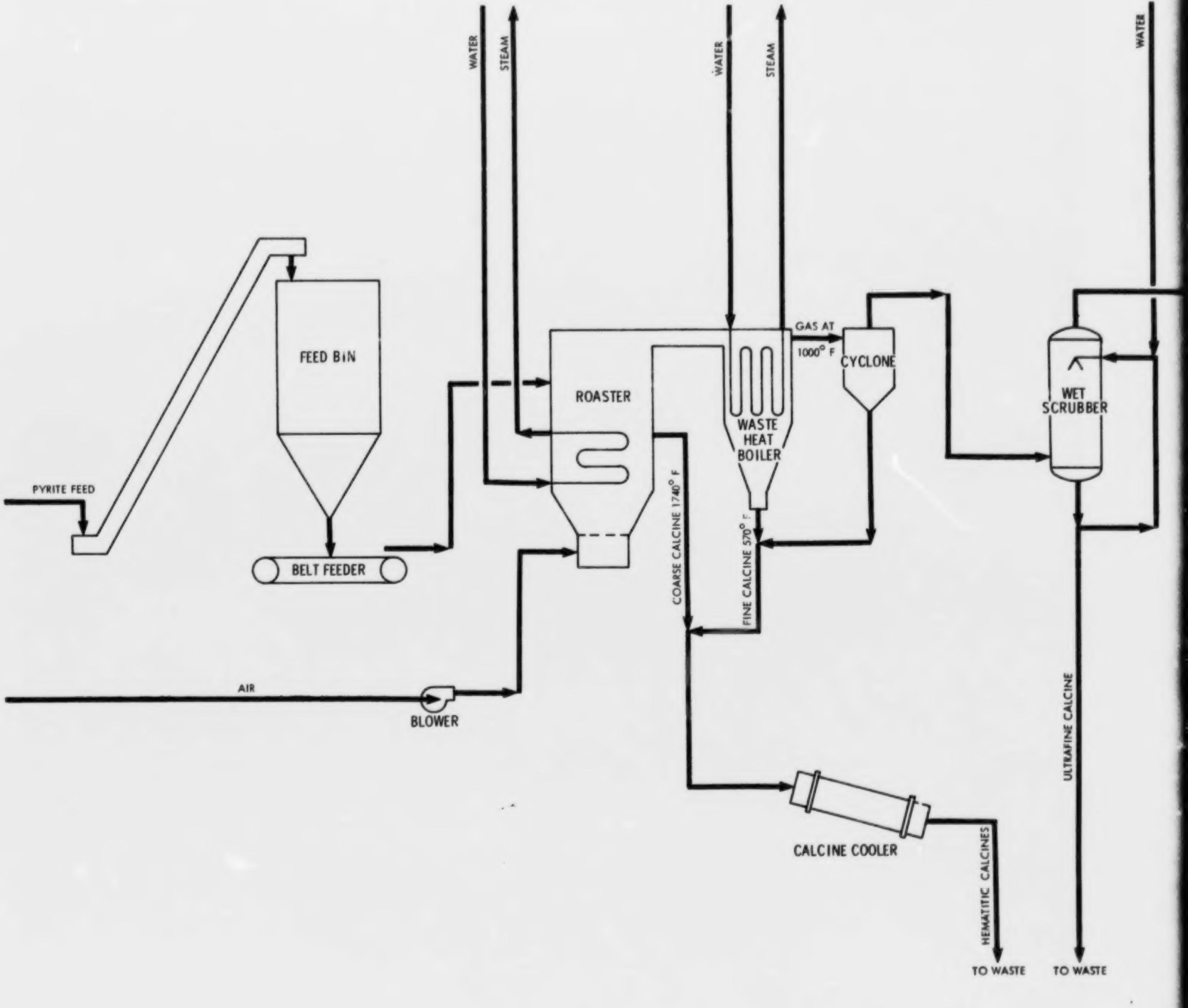
5.1.2 Adaptation to Pyrite-Coal Feed

A pyrite-coal mixture could be roasted in a fluid-bed, but laboratory work would be needed to determine the operating temperature for various ratios of coal to pyrite. Such a program would mostly concern the characteristics of the ash associated with the coal. To maintain fluidization, the pyrite-coal must be roasted below the incipient fusion temperature of the ash. Coking properties of the coal, due to volatiles, would probably not be important because of the high roasting temperature and nature of fluid-bed operation.

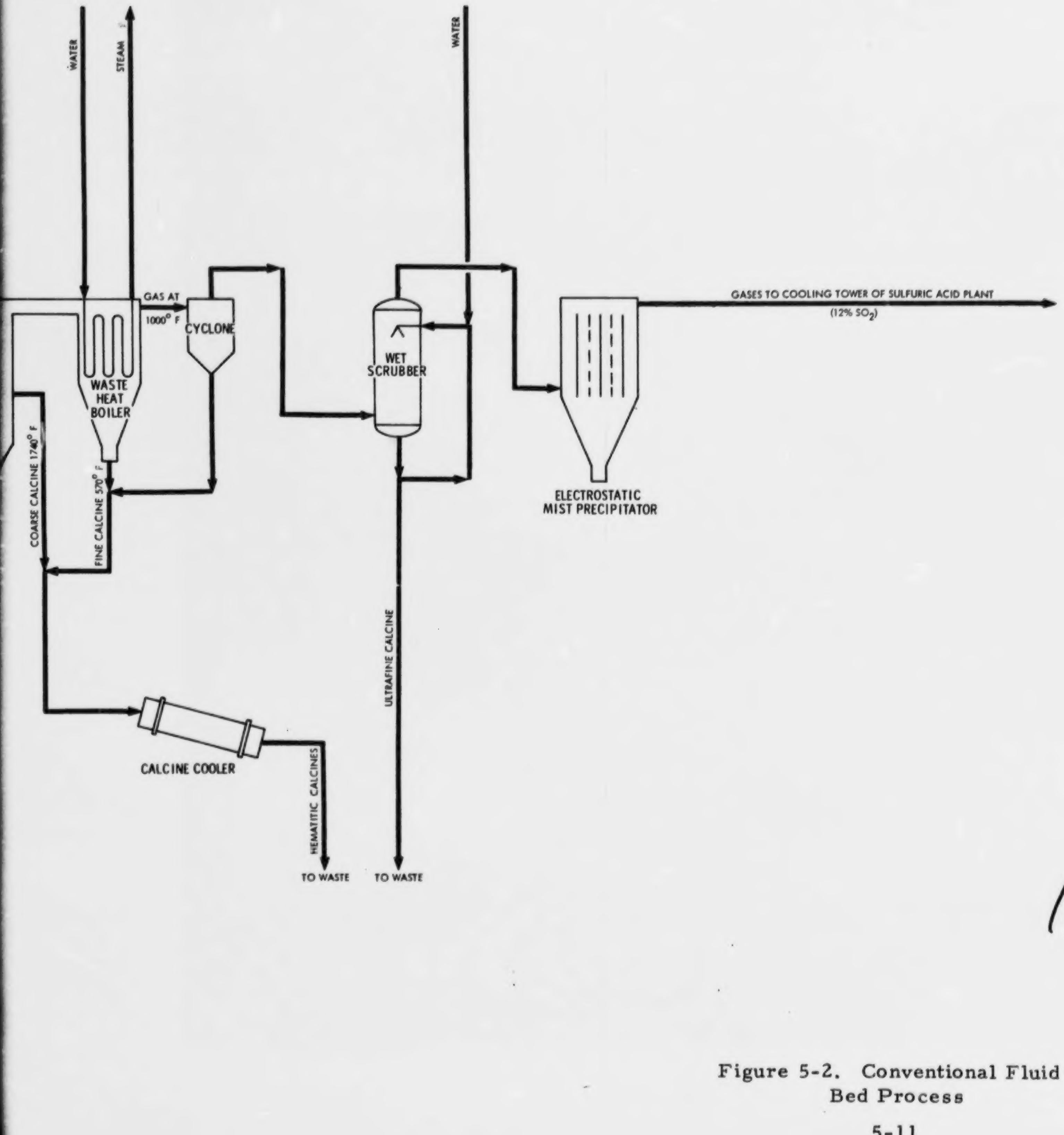
Roasting pyrite generates considerable heat and requires careful attention to control the temperature. This is done by adding water to the feed and injecting it as a slurry or by spraying water into the reactor. The presence of coal with pyrite would make temperature control more critical. Increased use of water to control temperature places an extra heat load on the gas scrubber that precedes the acid plant. The greater volume of water, furthermore, would result in more soluble loss of sulfur dioxide.

As the ratio of coal to pyrite increases, moreover, SO_2 concentration of the roaster gases decreases. Typical contact acid plants process a feed gas containing 5 to 13 percent SO_2 by volume. Conventional pyrite roasting can easily produce gas within this range. Off-gases from roasting mixtures of pyrite-coal would be more dilute, but would still be generally in this range down to mixtures as low as 50 percent pyrite.

Table 5-2 shows approximate SO_2 concentrations of gases from roasting pyrite-coal. It is assumed that the gases have been scrubbed then dried to 80 F saturation humidity and are ready to enter the contact acid plant; that enough excess air is present to oxidize SO_2 to SO_3; and that the nonpyrite fraction of the roaster feed is 50 percent coal and 50 percent inert mineral matter.


PYRITE FEED
FEED BIN
BELT FEEDER
AIR
BLOWER
WATER
STEAM
ROASTER
WATER
STEAM
GAS AT 1000° F
WASTE HEAT BOILER
CYCLONE
COARSE CALCINE 1740° F
FINE CALCINE 570° F
CALCINE COOLER
HEMATITIC CALCINES
TO WASTE
WATER
WET SCRUBBER
ULTRAFINE CALCINE
TO WASTE

A


WATER
STEAM
GAS AT
1000° F
CYCLONE
WASTE
HEAT
BOILER
WATER
WET
SCRUBBER
GASES TO COOLING TOWER OF SULFURIC ACID PLANT
(12% SO_2)
ELECTROSTATIC
MIST PRECIPITATOR
COARSE CALCINE 1740° F
FINE CALCINE 570° F
ULTRAFINE CALCINE
CALCINE COOLER
HEMATITIC CALCINES
TO WASTE
TO WASTE

B

Figure 5-2. Conventional Fluid-Bed Process

Table 5-2

SO_2 CONCENTRATIONS
OFF-GAS FROM PYRITE-COAL ROASTING

Percent Composition			Percent SO_2
Pyrite	Coal	Ash	In Off-Gas
100	0	0	11.3
90	5	5	9.8
80	10	10	8.4
70	15	15	7.2
60	20	20	5.9
50	25	25	4.8
40	30	30	3.7

Assumptions

1. Enough excess air is present to oxidize SO_2 to SO_3
2. Gas is saturated with water vapor at 80 F

These calculations show that roasting pyrite-coal mixtures will produce an off-gas suitable for a contact acid plant. With special design, it would be theoretically possible to roast a 40-percent pyrite (for 3.7 percent SO_2). Difficulties accumulate below this percentage of SO_2 in the off-gas. When SO_2 concentration falls below 3.5 percent, the acid plant is no longer in thermal balance, and heat must be added to the system. Below 3.5 percent SO_2, because more dilute gas requires larger equipment, capital and operating costs increase rapidly.

A further technical problem appears when the SO_2 concentration falls below 3 percent. The quantity of water brought into the contact acid system by humidity in the gas begins to dilute the acid product. This can only be overcome by refrigerating the gas to reduce humidity. This in turn increases costs.

Depending on the color of the acid produced, marketing problems may exist. Acid from smelters and from the Monsanto process is discolored due to organic carry-over. Pilot plant operation would be needed to evaluate the extent of discoloration. Although the market for dark sulfuric acid would be limited to fertilizer manufacture, this would not be a severe handicap since about half of U. S. acid production goes into this use.

Roasting pyrite-coal is one of the more attractive of the processes considered during this study; the capital investment is modest, and the technical problems do not appear imposing. However, since A. D. Little and Dorr-Oliver are examining this concept under separate contract with the Public Health Service, it will not be pursued further in this report.

5.2 OUTOKUMPU PROCESS

The Outokumpu process uses a flash smelting operation that removes the labile sulfur from the FeS_2, producing elemental sulfur and FeS matte. The remaining sulfur is recovered as acid. A coarse-grained iron oxide is also produced which Outokumpu claims is suitable for blast-furnace steelmaking. They also state that the nonferrous fraction of the ore is beneficiated to a degree that makes recovery of these values a possibility.

The Outokumpu Company's plant at Kokkola, Finland, is a full-scale commercial operation that produces approximately 90,000 tons per year of elemental sulfur, 290,000 tons per year of sulfuric acid, and 250,000 tons per year of iron oxide pellets. At the same time it feeds about 660,000 kwh per day of electricity into the power grid of Finland. The plant feed is a pyrite flotation concentrate.

A plant using this process has been proposed for the Brunswick Mining and Smelting Company near Belledune Point, N. B., Canada, by the Lummus Company, who is the licensor for the process. Although the plant was not constructed some engineering was done and some production estimates developed.

The feed to both the plant in Finland and the plant designed for Canada comes from pyrite ores that are about 38 percent sulfur and 35 percent iron. These ores contain some lead, zinc, copper, and silver that are partially separated from the pyrites by flotation. The flotation tailings, or pyrite concentrates, become the feed to the process.

The Brunswick plant was to convert 1.1 million tons per year of pyrite ore to 625,000 tons of iron pellets (67 percent Fe), 850,000 tons of sulfuric acid, and 250,000 tons of sulfur. The anticipated recovery of the nonferrous metals corresponding to 500,000 tons per year of pyrite concentrates is listed in Table 5-3.

Table 5-3

PROJECTED PRODUCTION OF BRUNSWICK PYRITE-BURNING PLANT

UNIT	PRODUCTS	QUANTITY*
Smelter	Sulfur	113,000
	Zinc, Lead	11,000
	Silver	40,000 oz
	Pyrrhotite to roasting plant	
	Steam to power plant	
Power Plant	Electricity	35 Mw
Roasting Plant	Metal oxides to chlorination	
	Sulfur dioxide to acid plant	
Acid Plant	Sulfuric acid	298,000
Chlorination Plant	Zinc	3,000
	Copper	2,000
	Magnetite to pelletizing plant	
Pelletizing Plant	Iron pellets	265,000

Based on 500,000 Tons/Yr Concentrated Feed

*Tons per year except where otherwise noted.

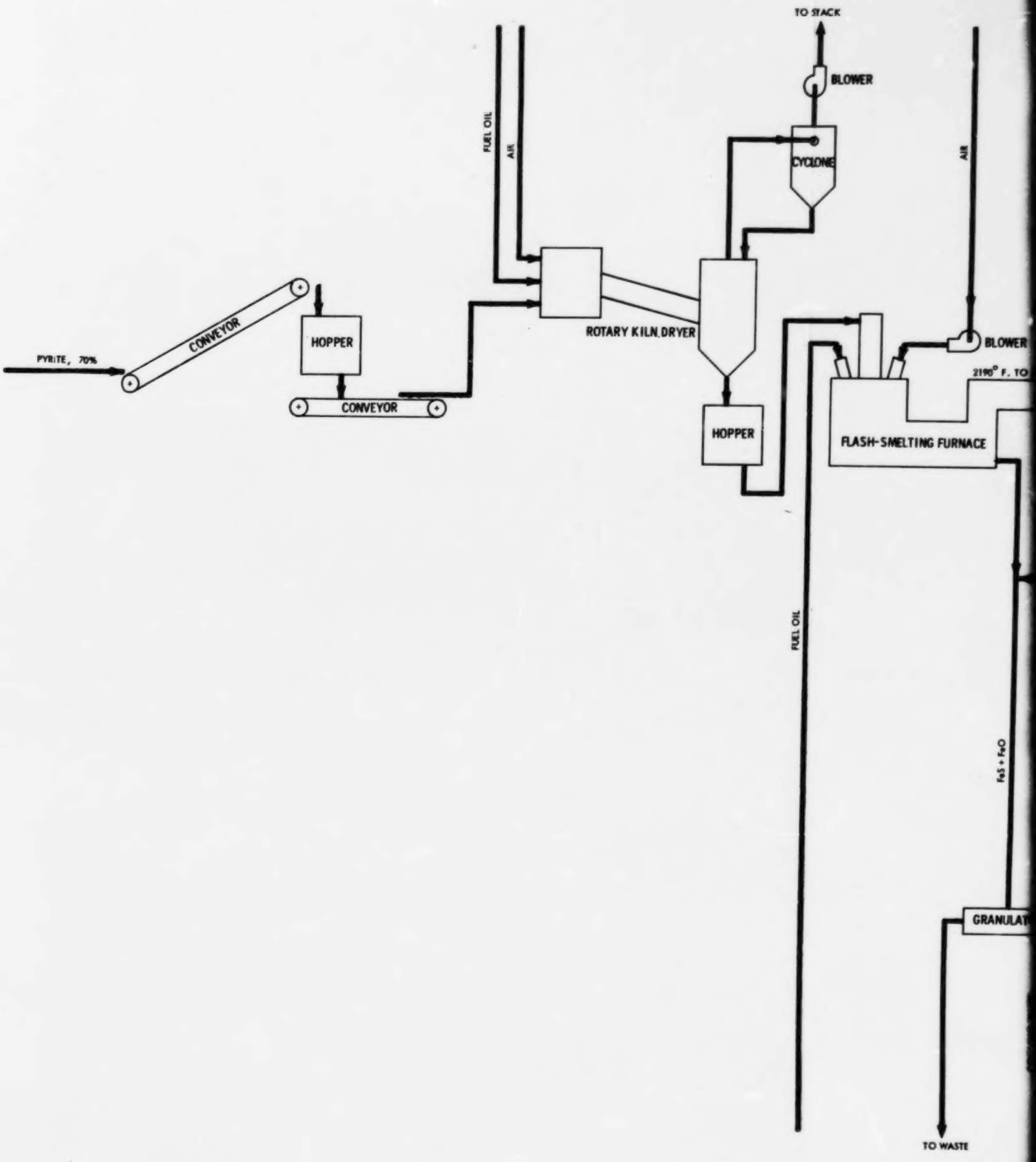
The major steps in the Outokumpu process are:

- Pyrite Drying: The pyrite ores are dried in a rotary kiln.
- Pyrite Smelting: Dry fine-grained pyrite is flash smelted with combustion gases at a temperature high enough to decompose the FeS_2 and yield its labile sulfur. The remaining molten FeS collects below in the reverbatory section of the furnace.
- Heat Recovery and Gas Cleaning: Sulfur-bearing hot gases from the reverbatory furnace are cooled in boilers to recover waste heat and are sent through an electrostatic precipitator to reduce solid carry-over.
- Recovery of Sulfur: Elemental sulfur is removed from the cleaned gases in two stages.
- Elimination of Arsenic: The sulfur is treated with milk of lime to remove the arsenic.
- Roasting: The liquid FeS matte is granulated in water, then roasted to give a coarse-grained iron-oxide product and SO_2-rich roasting gas. The sulfur dioxide is converted to sulfuric acid in a conventional contact plant.
- Generation of Electricity: Steam generated by the hot gases is used to generate electrical power.

5.2.1 Outokumpu Process (Finland)

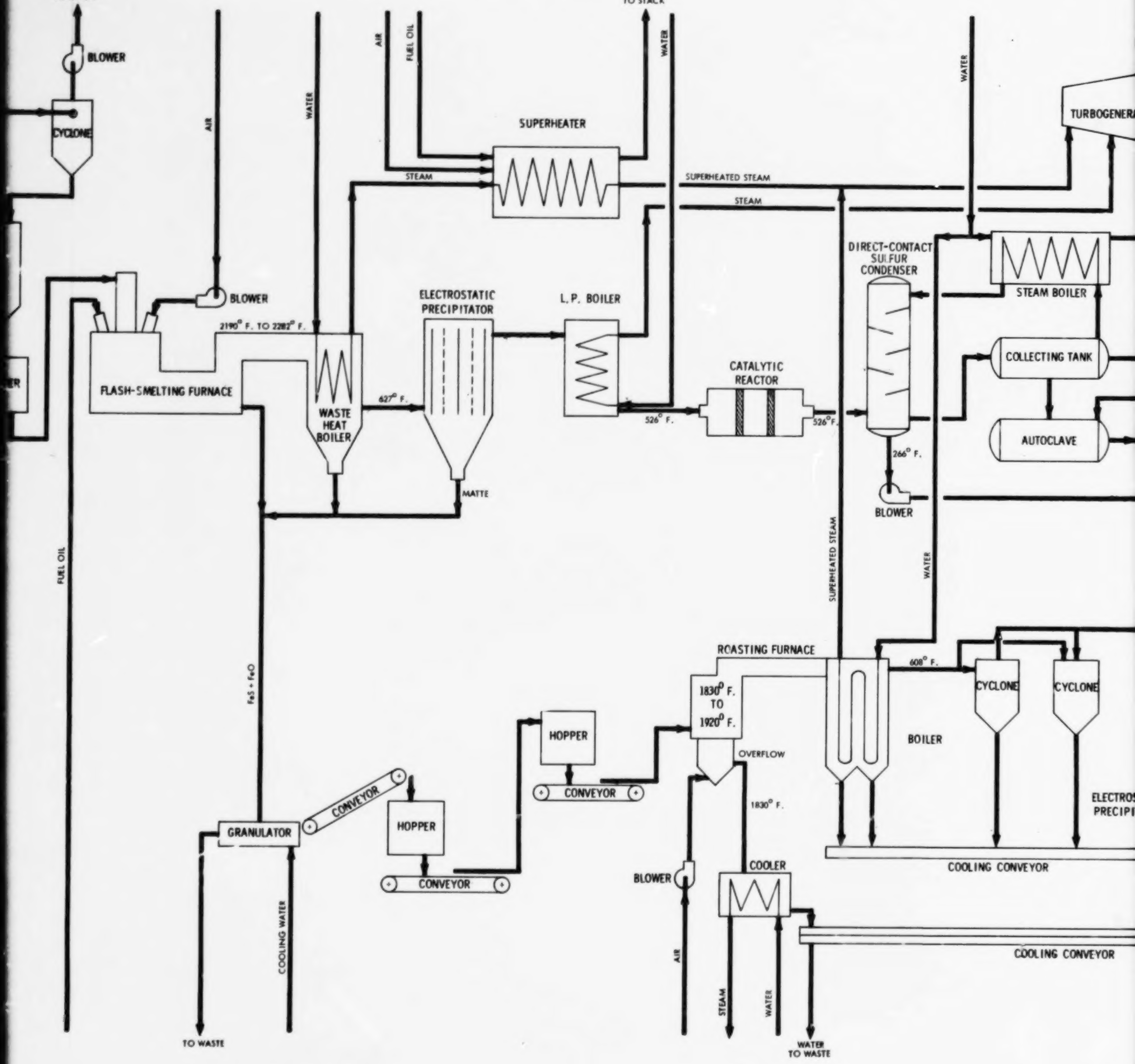
The process description and flowsheet which follow refer specifically to the process at the Outokumpu Plant in Finland. It is believed that the description approximately describes the process design for the New Brunswick plant, but excludes equipment related to the recovery of non-ferrous metals. Figure 5-3 shows the process flow.

- Drying: The first phase of the process dries the concentrates in a rotary kiln by means of hot combustion gases.
- Pyrite Smelting: The dry concentrate feeds into the upper part of the furnace and hot combustion gases flow into the same space from the oil burners. It is possible to



TO STACK
BLOWER
CYCLONE
FUEL OIL
AIR
AIR
PYRITE, 70%
CONVEYOR
HOPPER
CONVEYOR
ROTARY KILN DRYER
HOPPER
BLOWER
2190° F. TO
FLASH-SMELTING FURNACE
FUEL OIL
FeS + FeO
GRANULAT
TO WASTE

A


TO STACK
BLOWER
CYCLONE
AIR
WATER
AIR
FUEL OIL
TO STACK
WATER
SUPERHEATER
STEAM
SUPERHEATED STEAM
STEAM
WATER
TURBOGENERA
BLOWER
2190° F. TO 2282° F.
FLASH-SMELTING FURNACE
WASTE HEAT BOILER
627° F.
ELECTROSTATIC PRECIPITATOR
MATTE
L.P. BOILER
526° F.
CATALYTIC REACTOR
526° F.
DIRECT-CONTACT SULFUR CONDENSER
266° F.
BLOWER
STEAM BOILER
COLLECTING TANK
AUTOCLAVE
FUEL OIL
SUPERHEATED STEAM
WATER
FeS + FeO
ROASTING FURNACE
1830° F. TO 1920° F.
OVERFLOW
1830° F.
608° F.
BOILER
CYCLONE
CYCLONE
ELECTROS PRECIPI
HOPPER
CONVEYOR
HOPPER
CONVEYOR
CONVEYOR
GRANULATOR
COOLING WATER
TO WASTE
BLOWER
COOLER
AIR
STEAM
WATER
WATER TO WASTE
COOLING CONVEYOR
COOLING CONVEYOR

B

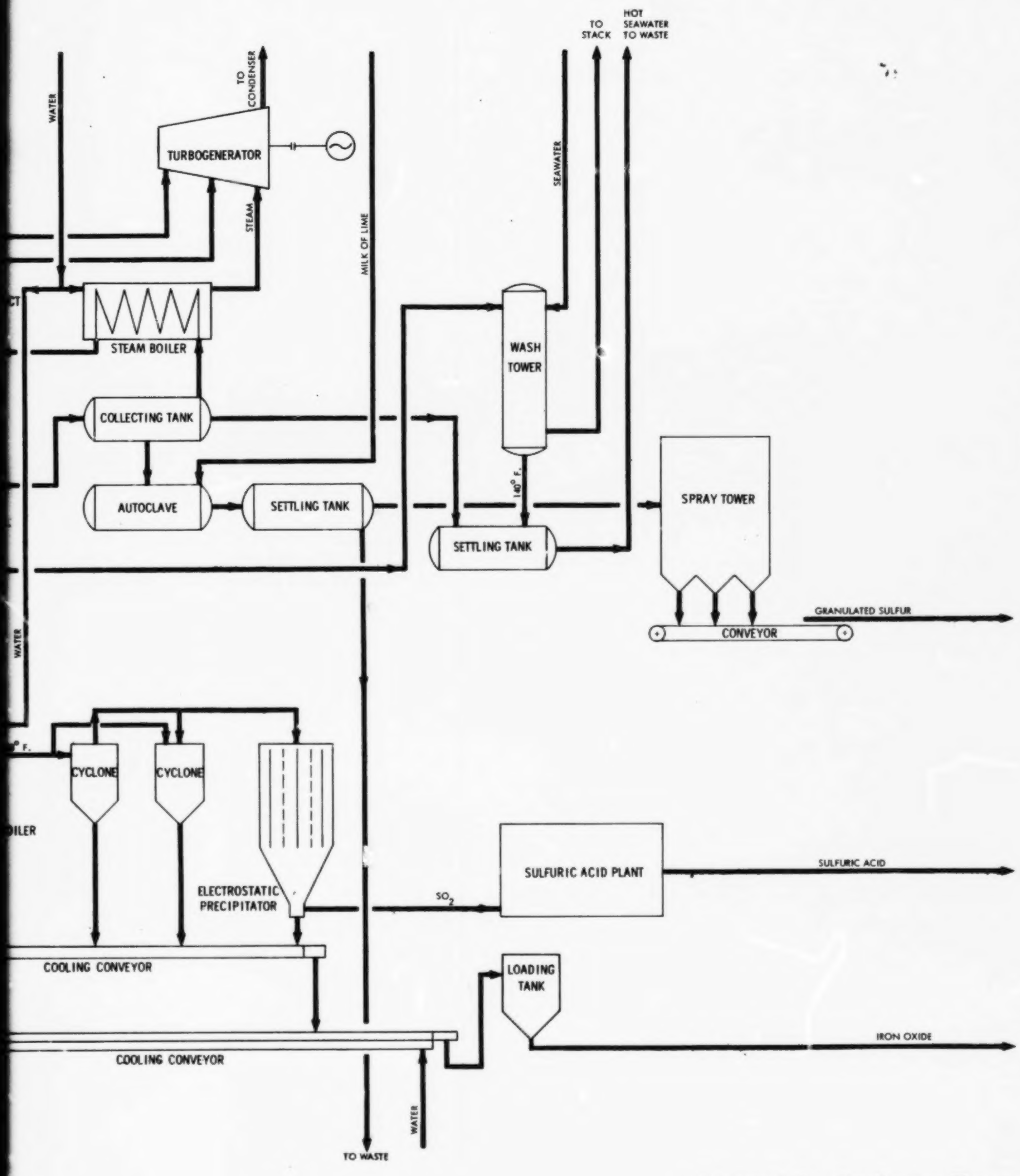

TO CONDENSER
WATER
TURBOGENERATOR
STEAM
MILK OF LIME
TO STACK
HOT SEAWATER TO WASTE
SEAWATER
STEAM BOILER
WASH TOWER
COLLECTING TANK
AUTOCLAVE
SETTLING TANK
140° F.
SETTLING TANK
SPRAY TOWER
GRANULATED SULFUR
CONVEYOR
WATER
CYCLONE
CYCLONE
ELECTROSTATIC PRECIPITATOR
SO2
SULFURIC ACID PLANT
SULFURIC ACID
COOLING CONVEYOR
LOADING TANK
IRON OXIDE
COOLING CONVEYOR
WATER
TO WASTE

Figure 5-3. Outokumpu Process

C

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burn heavy oil (Bunker C) in special high-efficiency burners without soot formation. The hot combustion gases flow at a controlled rate so they suspend the concentrate particles. Heat transfer, thermal decomposition, and chemical reactions take place in this suspension. In the reverbatory section, molten particles separate to form a liquid bath composed of iron sulfide (FeS) in which some iron oxide has dissolved. The gangue from the concentrate forms slag which separates on the surface of the molten FeS matte. The liquid FeS is tapped from the furnace and granulated in water. Its grain size is about 4 mm and smaller. Very fine grain sizes (i.e., dust-like material) do not form during granulation. The chemical analysis of the FeS matte is:

Fe = 63.2 percent S = 27.0 percent

- Heat Recovery and Gas Cleaning: At about 2280 F, the furnace gases contain not only combustion gases but also sulfur vapor, SO_2, H_2S, and some flue dust. These gases flow through a high-pressure steam boiler. The temperature of the tubes is so high that no condensation of elemental sulfur on the tubes takes place. The first part of the boiler consists of a spacious radiation chamber with tube-lined walls in which gases are cooled down to about 1290 F. In the extension of the radiation chamber (the convection section), the gases are further cooled to about 660 to 610 F. Steam from the high-pressure boiler has a pressure of 1000 psig and a temperature of 545 F. This steam is superheated separately and used together with all other steam generated at various points in the system for the production of electrical energy. The cooled gases are led into an electrostatic precipitator with a sufficiently high separation efficiency to give a low-ash content in the sulfur. The separated dust consists of fine-grained particles of iron sulfide and oxide plus zinc sulfide, which volatizes during the smelting process.

- Recovery of Sulfur: The cleaned gases flow into a low-pressure steam boiler (50 psig, 302 F). Since the temperature of the tubes is below the dew point of the sulfur vapor, elemental sulfur condenses and collects at the bottom of the boiler.

  The gas temperature is further lowered to a point favorable for the catalytic reaction of SO_2 and H_2S to form elemental sulfur. The catalyst chamber is designed to allow the reaction to approach equilibrium.

Sulfur is separated from the gases by contact with circulating molten sulfur in a tower. Fine sulfur drops cool the gas and sulfur vapor condenses on the surface of these drops. Molten sulfur is collected in a tank, from which a part of it is piped through cooling elements into the gas cooling tower. The heat from the circulating sulfur is used for generating low-pressure steam (10 psia). After the gases have been washed by molten sulfur, they are led into a wash tower to be further cooled and washed by seawater. The temperature falls from 226 to 140 F, and a further small amount of sulfur is recovered in a settling tank into which water is led from the wash tower. From this wash tower, the gases are led into the stack.

- Elimination of Arsenic: Molten sulfur produced in the low-pressure steam boiler and in the sulfur washer proper is of high purity except that it contains the greater part of the arsenic in the concentrate. This arsenic content is higher than desirable for a marketable product and therefore further cleaning is carried out in an autoclave. The molten sulfur flows through a horizontal autoclave where it contacts milk of lime. The cleaned sulfur is then piped into a granulation and storing tower. The sulfur is granulated by spraying it into air through a special nozzle; the sulfur drops are cooled and solidified.

  The average monthly analysis of the sulfur product shows the following impurities:

  As = 0.005 percent Se = 0.004 percent Ash = 0.015 percent

- Roasting: The temperature for roasting FeS matte is 1830 to 1920 F. This is 100 to 200 F higher than for common pyrite roasting. The greater part of the roasted product, 80 to 90 percent, overflows from the furnace bed at 1830 F temperature and is cooled in water-cooled screw and Redler conveyors. The heat content is recovered and used for preheating boiler feedwater. The fine fraction which is carried over with the gases is recovered in boilers, cyclones, and electrostatic precipitators.

  A typical chemical analysis of the roaster calcines is:

  Fe = 67 percent S = 0.5 percent Cu = 0.06 percent

  Zn = 0.08 percent As = 0.007 percent

With regard to impurities, copper goes into FeS matte and into the calcines. Zinc is concentrated in the flue dust of the smelting furnace.

The furnace gases are cooled in steam boilers to about 610 F and are cleaned in cyclones and then in electrostatic precipitators. The gases then pass into an adjacent sulfuric acid plant.

Hot cinders from the furnace are cooled in a special screw conveyor, then further cooled in Redler conveyors.

- Generation of Electric Power: Steam generated by the smelting furnace gases has a pressure of 1000 psig and is heated to 932 F in a separate superheating boiler connected with the power plant. Low-pressure steam is also recovered from the process. The roasting plant produces a superheated steam at 490 psig. All steam is used for generating electric power.

The oil-burning step aims at complete combustion with nearly the stoichiometric quantity of air. The combustion gases, about 3270 F, consist mainly of CO_2, N_2, and H_2O. In addition to these components, some CO, H_2, and O_2 are present as the result of dissociation, the amount depending on the amount of excess air used.

However, the molten pyrrhotite particles from decomposition of pyrite are not pure FeS, but contain some iron oxide (FeO). Likewise, the sulfur in the gas phase does not occur only as S_2 vapor, but the gas also contains SO_2, H_2S, and some COS. Thus, there occur in the molten and in the gas phase the following primary and secondary components:

	Primary Components	Secondary Components
Gas	CO_2, H_2O, S_2, N_2	SO_2, H_2S, CO, H_2, COS
Liquid	FeS	FeO

The process temperature is regulated so that the resulting components tend to give as high a rate of sulfur recovery as possible. Although in principle the combustion of the oil is stoichiometric, the ratio of $H_2S:SO_2$ in the gas can be influenced by a relatively small change in the excess air.

When leaving the furnace, the gases contain sulfur, mainly in the form of sulfur vapor, and partly in the form of sulfur-bearing gaseous components. During cooling, changes in the equilibrium conditions take place so that the content of elemental sulfur increases. During the first stage of gas cooling (from 2280 to 1110 F), CO and H_2 react with SO_2, forming S_2 and H_2S. During the second stage of cooling (from 1110 to 570 F), the kinetic possibilities for these reactions decrease. Then an increase in elemental sulfur takes place according to the following reactions:

$$2\ H_2S + SO_2 \rightarrow 2\ H_2O + 3\ S$$
$$2\ COS + SO_2 \rightarrow 2\ CO_2 + 3\ S$$

This reaction is promoted by an aluminum-based catalyst; the optimum temperature is around 536 F.

5.2.2 Adaptation to Pyrite-Coal Feed

The Outokumpu process recovers about half the sulfur in marketable elemental form. This process could be adapted to feed coal-pyrite mixtures, but there are many reasons to believe that the economics would be quite poor, one being that the process depends heavily on producing a marketable iron oxide.

It is interesting to note that coal-pyrite supplies its own fuel requirement. However, a special burner and feed system would have to be developed for the flash furnace that would operate with coal. It would have to provide complete combustion with very little excess air, so that the sulfur recovery system would be similar to that described for conventional pyrite feed.

The most difficult problem would arise in the formation of slag and iron matte. After the FeS matte is recovered, it can be roasted to give SO_2 for an acid plant. However, the ash in the coal must be slagged with iron to form a flux having a critical composition that gives the required fluidity. This adjustment would be difficult and decrease iron yield.

The economics of the Outokumpu process are not favorable with conventional pyrite ore. The process could probably be adapted to coal pyrite, but the modified plant would be more expensive to build and operate than would a plant for conventional pyrite ore and the yield of products would be lower. In conclusion it appears that the process was originally designed for a specific ore in Finland. Further development work is needed to adapt the process to other ores.

5.3 ORKLA PROCESS

The Orkla process smelts cupriferous pyrite in a specially designed blast furnace to produce elemental sulfur and gaseous sulfur compounds. Iron, copper, and precious metals report in a matte. Waste heat boilers, steam turbines, and generators recover heat of reaction and produce enough electricity to run the plant. The Orkla process is limited to a lump-pyrite feed.

The Orkla process originated in Norway around 1928. The Norwegian plant now produces 115,000 tons per year of sulfur. Other Orkla plants have been licensed and built in Spain, Portugal, and Hungary. In Spain, the Orkla plant of the Rio Tinto Mines produced about 30,000 tons per year of elemental sulfur during 1960, 1961 and 1962. Production was gradually cut, and in 1965 it was about 14,000 tons. The major process steps are:

- Pyrite Smelting: Lump pyrite is smelted in a blast furnace. Elemental sulfur and gaseous sulfur compounds leave the top of the furnace. A low-grade copper matte is separated later by resmelting the slag-matte.
- Sulfur Recovery: Elemental sulfur leaving the furnace condenses in waste heat boilers. Gaseous sulfur compounds are catalytically converted to elemental sulfur, condensed, and recovered.

5.3.1 Orkla Process (Norway)

Figure 5-4 shows the process flow scheme. Table 5-4 gives a typical analysis of a Norwegian ore feed to the Orkla process.

Table 5-4

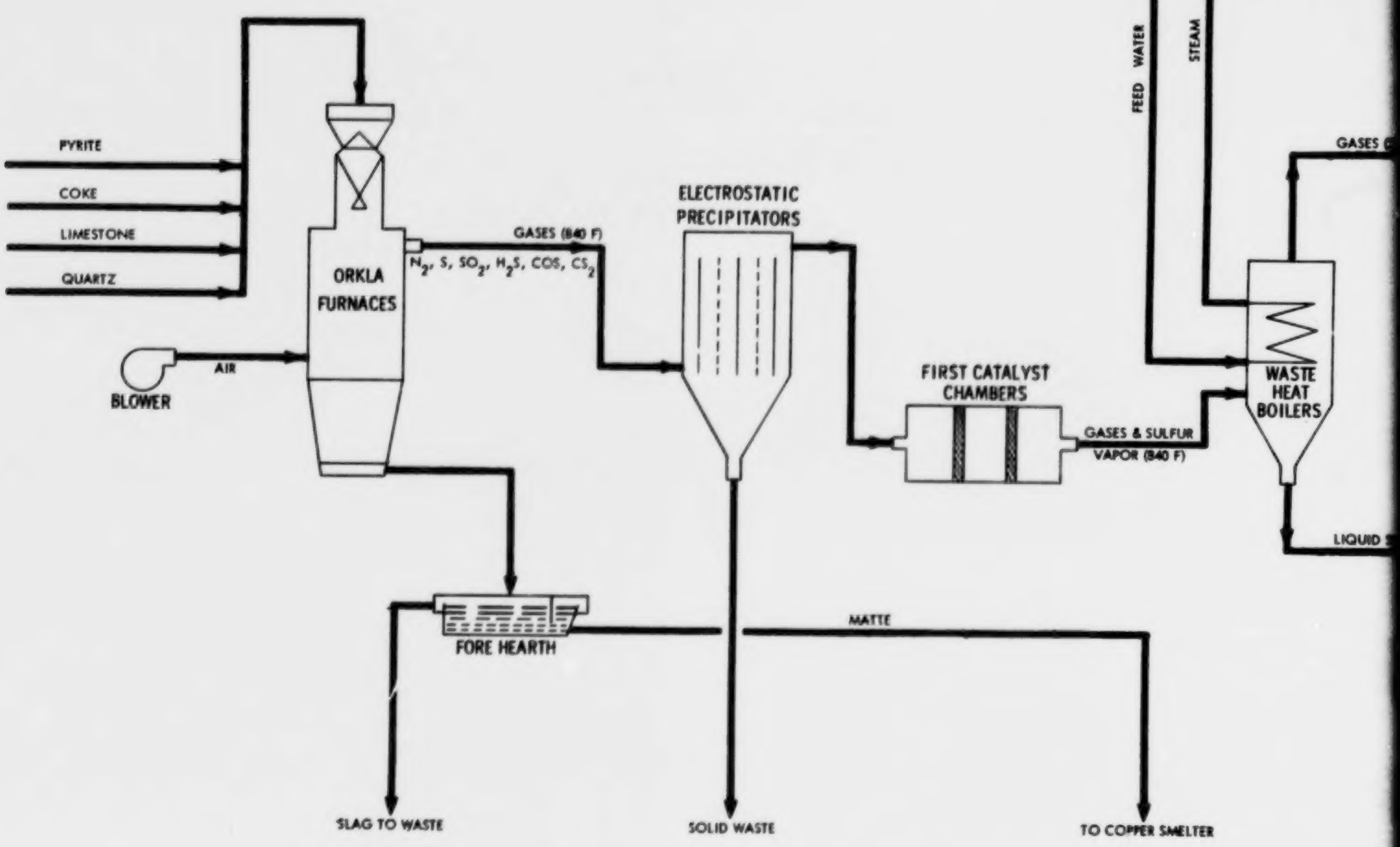
NORWEGIAN ORE ANALYSIS

Element	Percent
Sulfur	41.0
Copper	1.8 - 2.5
Iron	38.0
Zinc	1.9
Cobalt	0.1
Arsenic	0.05
Silver	20 g/ton
Gold	0.2 g/ton

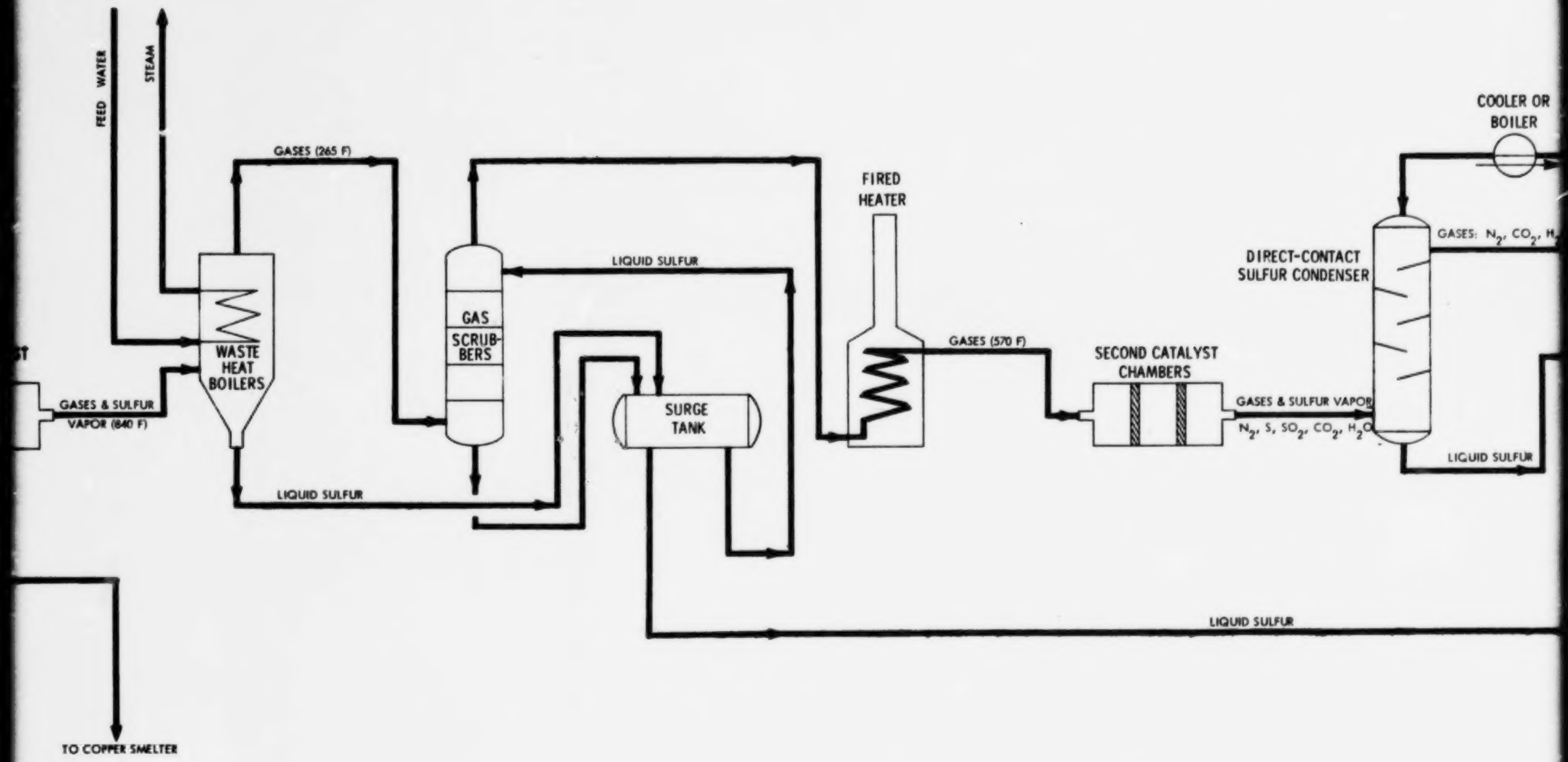
- Smelting: The Orkla furnace resembles an ordinary blast furnace that smelts sulfide ores to matte, except that it has a closed top like an iron blast furnace. Lump pyrite is mixed with coke, silica, and limestone, and charged into the top of the furnace. Air is blown in through tuyeres.

  The labile sulfur distills off in the upper part of the furnace and does not burn to SO_2, since no oxygen is present.

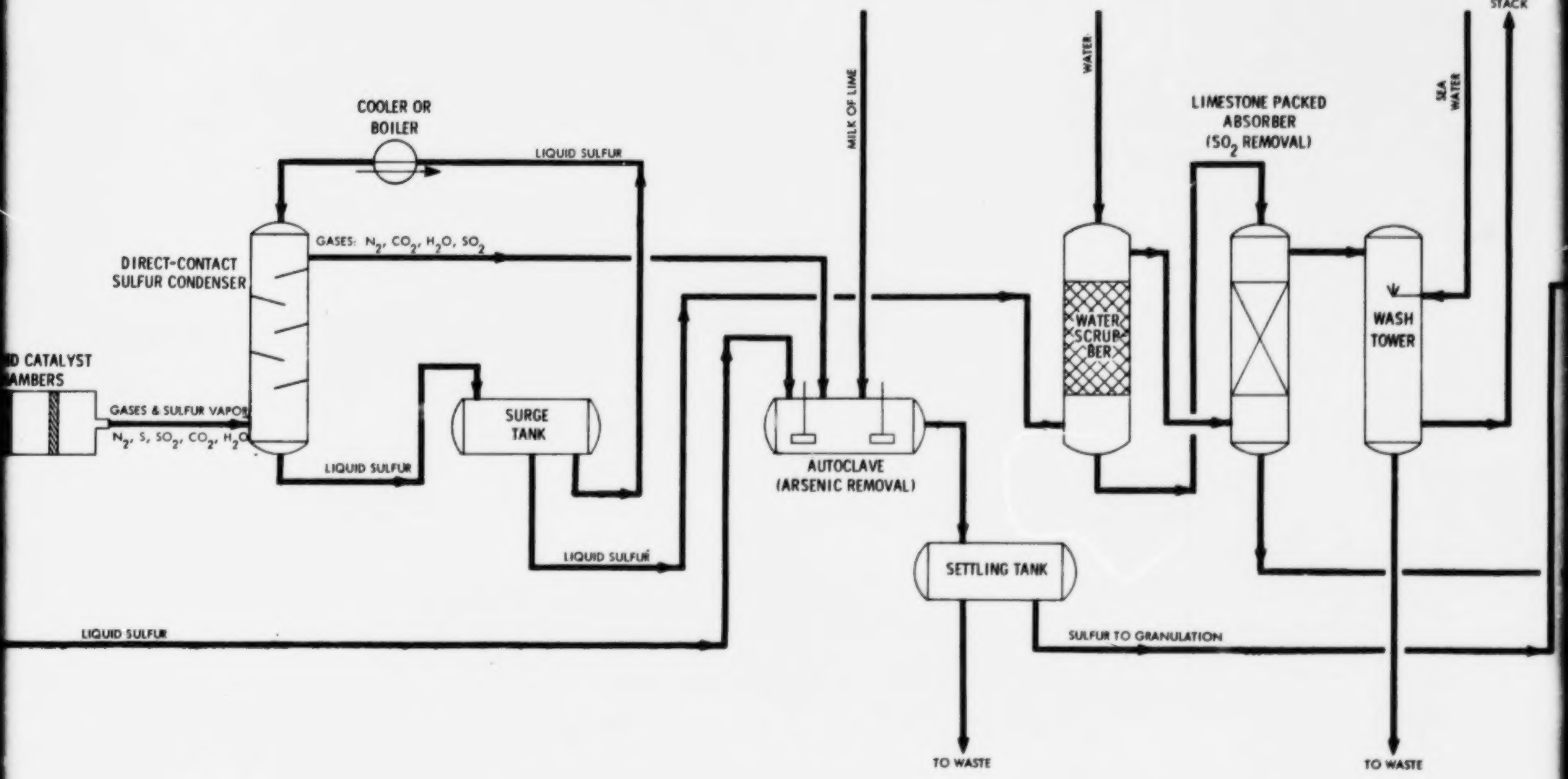
$$7\ FeS_2 \rightarrow Fe_{.7}S_8 + 3\ S_2$$


PYRITE
COKE
LIMESTONE
QUARTZ
ORKLA FURNACES
AIR
BLOWER
GASES (840 F)
N_2, S, SO_2, H_2S, COS, CS_2
ELECTROSTATIC PRECIPITATORS
FIRST CATALYST CHAMBERS
GASES & SULFUR VAPOR (840 F)
FEED WATER
STEAM
GASES
WASTE HEAT BOILERS
LIQUID
FORE HEARTH
MATTE
SLAG TO WASTE
SOLID WASTE
TO COPPER SMELTER

A

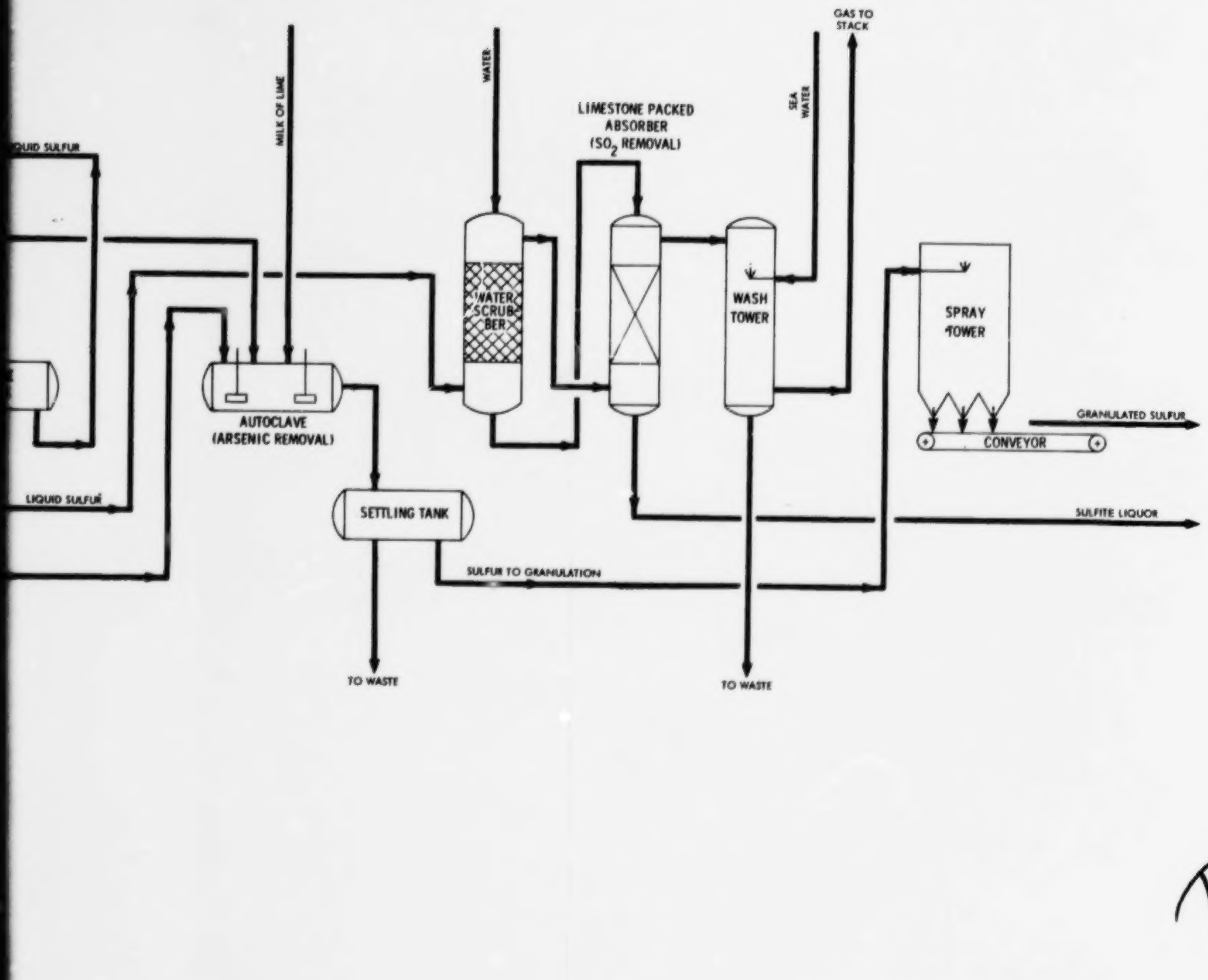

FEED WATER
STEAM
ST
GASES & SULFUR
VAPOR (840 F)
WASTE
HEAT
BOILERS
GASES (265 F)
LIQUID SULFUR
GAS
SCRUB-
BERS
LIQUID SULFUR
SURGE
TANK
FIRED
HEATER
GASES (570 F)
SECOND CATALYST
CHAMBERS
GASES & SULFUR VAPOR
N2, S, SO2, CO2, H2O
DIRECT-CONTACT
SULFUR CONDENSER
COOLER OR
BOILER
GASES: N2, CO2, H
LIQUID SULFUR
LIQUID SULFUR
TO COPPER SMELTER

B


ID CATALYST
AMBERS
GASES & SULFUR VAPOR
N_2, S, SO_2, CO_2, H_2O
DIRECT-CONTACT SULFUR CONDENSER
COOLER OR BOILER
LIQUID SULFUR
GASES: N_2, CO_2, H_2O, SO_2
LIQUID SULFUR
SURGE TANK
LIQUID SULFUR
LIQUID SULFUR
MILK OF LIME
AUTOCLAVE (ARSENIC REMOVAL)
SETTLING TANK
TO WASTE
SULFUR TO GRANULATION
WATER
WATER SCRUBBER
LIMESTONE PACKED ABSORBER (SO_2 REMOVAL)
SEA WATER
WASH TOWER
GAS TO STACK
TO WASTE

Fi

e


QUID SULFUR
LIQUID SULFUR
MILK OF LIME
AUTOCLAVE
(ARSENIC REMOVAL)
SETTLING TANK
TO WASTE
SULFUR TO GRANULATION
WATER
WATER SCRUBBER
LIMESTONE PACKED
ABSORBER
(SO_2 REMOVAL)
SEA WATER
GAS TO STACK
WASH TOWER
TO WASTE
SPRAY TOWER
CONVEYOR
GRANULATED SULFUR
SULFITE LIQUOR

D

Figure 5-4. Orkla Process

The resulting Fe_7S_8 descends to the smelting zone and oxidizes:

$$2\ Fe_7S_8 + 23\ O_2 \rightarrow 14\ FeO + 16\ SO_2$$

Most of the SO_2 is reduced to elemental sulfur by coke:

$$SO_2 + C \rightarrow CO_2 + 1/2\ S_2$$

The amount of coke fed to the furnace is regulated so that some SO_2 remains in the gases leaving the furnace. This is done so that there will be enough SO_2 to react with the other sulfur compounds (COS, H_2S, CS_2) in later process steps.

In the smelting zone, the copper plus some FeS form a CuS-FeS matte running about 12-percent copper. FeO reacts with silica and limestone to form a slag containing most of the iron as silicates.

- Sulfur Recovery: Gases leave the top of the furnace at about 840 F. They consist of N_2, CO_2, SO_2, H_2S, COS, CS_2, and S_2, with a total sulfur content of about 300 grams per cubic meter. A hot electrostatic precipitator cleans the furnace gases before they pass through an aluminum-based catalyst which promotes the reactions:

$$SO_2 + 2\ H_2S \rightarrow 2\ H_2O + 3S$$
$$SO_2 + CS_2 \rightarrow CO_2 + 3\ S$$
$$SO_2 + 2\ COS \rightarrow 2\ CO_2 + 3\ S$$

The reacted gases then pass through a waste heat boiler, where they condense sulfur as they cool from 840 to 265 F. Gases from the boiler are reheated to about 570 F in a fired heater and pass through a second stage of catalysis to optimize the conversion to elemental sulfur. This sulfur is condensed in a cooling tower by direct contact with circulating molten sulfur.

- Sulfur Purification: The crude sulfur distilled directly from the pyrite, or formed by reactions, contains some arsenic, and possibly some ash. Arsenic is removed by circulating milk of lime through molten sulfur in autoclaves at about 260 F and 50 psi. The probable reaction is:

$$4\ Ca(OH)_2 + As_2S_5 + 2\ S_2 \rightarrow Ca_3As_2S_8 + CaSO_4 + 4\ H_2O$$

If the sulfur contains ash, this is removed by passing through steam-heated filter presses. In a final step, the sulfur product is granulated in a spray tower.

- Gas Cleaning: Gases from the second-stage catalysis contain N_2, CO_2, H_2O, and a small amount of SO_2. To reduce air pollution, SO_2 is absorbed in water scrubbers and reacted with limestone:

  $$2\ SO_2 + H_2O + CaCO_3 \rightarrow Ca(HSO_3)_2 + CO_2$$

  The reaction product, sulfite cooking liquor, may have a market in pulp and paper industries. Gases are given a final spraying with seawater before discharging to a stack.

- Copper Recovery: The copper from the smelting zone forms a CuS-FeS matte with the composition:

Cu	11.4%
Fe	54.0%
S	24.0%

  The Norwegian plant resmelts this matte with coke, silica, and limestone to upgrade it to about 40 percent. Then conventional copper converting and electrowinning produce refined copper.

  The slag, separated from the above matte in a forehearth, consists of about 80 percent $2(FeO)\cdot SiO_2$ and 15 percent $CaO\cdot FeO\cdot 2(SiO_2)$ along with small amounts of S, Zn, and Al_2O_3. The iron content of the slag is about 42 percent.

An inherent disadvantage of the Orkla furnace is that coke burns inefficiently. The process aims to produce as much SO_2 as possible at the focus and to have this travel up the column in such a way that it is almost completely reduced to sulfur before leaving the furnace. This reduction probably takes place in three stages:

$$(1)\quad SO_2 + C \rightarrow CO_2 + 1/2\ S_2$$
$$(2)\quad CO_2 + C \rightarrow 2\ CO$$
$$(3)\quad SO_2 + 2\ CO \rightarrow 1/2\ S_2 + 2\ CO_2$$

Unfortunately, the reduction of SO_2 by solid carbon (equation 1) only takes place with reasonable speed at about 2200 F. This high temperature zone in the furnace is too short to give efficient reduction. Some coke passes unburned down to the focus where it burns in air to CO_2. In this case, reduction of SO_2 takes place according to the reactions:

$$(4)\quad C + O_2 \rightarrow CO_2$$

$$(5)\quad CO_2 + C \rightarrow 2\,CO$$

$$(6)\quad SO_2 + 2\,CO \rightarrow 1/2\,S_2 + 2\,CO_2$$

Equations 4, 5, and 6 use twice as much coke as equations 1, 2, and 3. Actually, SO_2 is reduced by both routes. Figures from Rio Tinto show that about two thirds of the SO_2 reduce efficiently according to equations 1, 2, and 3, while the other one-third reduces according to 4, 5, and 6.

Another disadvantage results from the inefficient use of oxygen. The carbon reaching the focus consumes oxygen which would otherwise oxidize FeS and produce SO_2. The unreacted FeS lowers the copper content of the matte.

The Orkla plant in Norway successfully converts the gaseous sulfur compounds (H_2S, COS, CS_2) to elemental sulfur by catalysis. This is perhaps the chief reason why the plant can recover as much as 85 percent of the sulfur in the pyrite.

Catalysis has not been used at Rio Tinto because the ores contain considerable arsenic: 0.7 percent in Rio Tinto as opposed to 0.05 percent in Norway. As a result, the Orkla process at Rio Tinto recovers only about 55 percent of the sulfur in the pyrite, the major loss occurring in gases released to atmosphere.

A high arsenic content causes sulfur to condense at a higher temperature than with arsenic-free gases. The liquid sulfur condensed, containing high arsenic, is a viscous, sticky material that soon blankets the catalyst.

5.3.2 Adaptation to Pyrite-Coal Feed

The chemistry of the Orkla process would apply to pyrite-coal mixtures, but the technology would not be adaptable without major changes. A pyrite-coal mixture would have to be briquetted to be suitable feed to this process. Some attempts have been made to briquette conventional pyrite fines but apparently this has not been done economically.

The Orkla process is the only commercial-scale process that recovers most of the sulfur value in elemental form. This would give it a much stronger marketing position than processes that produce SO_2. The economics of the Orkla process, however, depend heavily on recovery of copper from the matte leaving the bottom of the furnace. At the present time, the plant at Rio Tinto is shut down, probably because their ore is running low on copper.

As with Outokumpu, the Orkla process is marginal economically when operating with conventional pyrite ores. The furnaces are expensive to build and operate. Some of the high operating cost is due to the requirement for coke. Pyrite-coal would eliminate this requirement, but the extra cost of briquetting and lack of a copper product would not make the economics attractive.

Section 6

FEASIBILITY CONSIDERATIONS AND PRELIMINARY ECONOMICS

Table 6-1 summarizes the economics for five of the seven processes described in the technology reviews. Not included are the Orkla and the preferential oxidation processes. These two were discarded for technical reasons. Cost data shown in the table include capital investments, total yearly costs, credits for products, and the costs of selling the products and transporting them to market. The change in clean coal price results from the net cost of processing the refuse. These preliminary economics were developed from published information on capital costs and material requirements.

The basis and criteria for estimating the economics are listed in paragraph 6.2. Assumptions regarding size and logistics of the operations are:

- A single mine produces 2,500,000 tons per year ROM coal that can be cleaned to a total sulfur content of 1.0 percent.
- The coal-preparation plant processes the ROM coal to produce 1,870,000 tons per year of clean coal and 630,000 tons per year of refuse.
- This quantity of clean coal is enough to feed a 1000 Mw(e) electric power plant. The fuel value in the refuse would feed a 250 Mw(e) plant.
- Processes are located adjacent to the coal preparation plant or the power plant.
- The market for by-products is within 200 miles.

Table 6-1

COST EVALUATION SUMMARY

	Combustion - SO_2 Removal	Fluid-bed Process	Outokumpu Process	Carbonization - Desulfurization	Solvent Extraction
	(figures in thousands except *)				
Capital Investment, $	8,300	2,400(1)	5,000	14,000	19,500
Total Annual Cost, $/Year	2,201	1,077	1,695	4,006	7,420
Credits for Products (FOB Plant), $/Year	1,160	826	867	2,432	4,368
Credits for Fuel Value, $/Year	1,230	36	60	—	—
Net Annual Cost, $/Year	-189	215	768	1,574	3.052
Change in Clean Coal Price,(2) $/ton	-0.10	+0.12	+0.41	+0.84	+1.63
Cost of Transporting Products to Market, $/Year	1,275	568	418	1,609	1,407
Selling Cost of Products, $/Year	100	58	61	160	278
Total Cost of Products to Customer at Market, $/Year	2,535	1,452	1,346	4,201	6,053

(1) Facilities include waste heat boilers and acid plant.

(2) Positive values indicate that cost of process exceeds credits from fuel value and products, which results in an increase in clean coal price. Negative values indicate that credits exceed costs, which results in a decrease in clean coal price.

6.1 COMPARISON OF PROCESSES

Factors of significance in technologic and economic comparison of the processes considered are tabulated in Table 6-2. Table 6-2 shows that the most economic method for processing coal refuse is by burning the refuse in a boiler, then recovering the sulfur as acid. Estimated revenue from sale of the acid slightly exceeds the cost of recovery. This scheme depends upon the willingness of an electric utility to install a power generator near the coal preparation plant and to operate a specially designed system on high-ash, high-sulfur fuel. The incentive for the utility would be the price charged for the fuel. It should be pointed out that processes of this type are in the pilot-plant stage and that acid concentration and quality are lower than produced by existing commercial methods.

The most attractive alternative to combustion-SO_2 recovery is to concentrate the pyrite content of the coal refuse and then to roast the concentrate in a fluid bed like conventional pyrite ore. A 90-percent pyrite concentrate would contain as much sulfur as most natural pyrite. Factors important to the economics are:

- An adequate market for the acid
- Adequate and economic recovery of a concentrate for feed to the roaster

The Outokumpu process, also operating on a 90-percent pyrite concentrate, ranks third in economy. This process suffers a penalty for the scale of operation, being originally intended for much larger production. Economy depends upon marketing iron ore as well as sulfur value. Such a process could only be considered for very large-scale coal mining, perhaps using the concentrates from several mines. Or, the feed to the coal preparation plant could be supplemented by existing waste piles.

Table 6-2

PROCESS EVALUATION FACTORS

	MARKETABLE BY-PRODUCTS	ADVANTAGES	DISADVANTAGES	ECONOMIC RESULT
COMBUSTION-SO_2 REMOVAL	Sulfuric acid Electric power	Requires no pyrite concentration; uses entire refuse stream Recovers both inorganic and organic sulfur Based relatively simple, well-known technical principles Products usually marketable	Sulfuric acid product is slightly discolored and limited to about 78 percent concentration Market radius is limited by shipping cost of acid (usually about 200 miles maximum) Final solution is required to select economical corrosion resistant materials of construction.	Reduces clean coal cost $0.10/ton Most economical process
CARBONIZATION-DESULFURIZATION	Sulfur Pipeline gas Char	Requires no pyrite concentration; uses entire refuse stream Recovers sulfur in elemental form	Complex, multioperation process with many technical problems to solve High-ash char may not be marketable; this product may retain sulfur	Adds $0.84/ton to clean coal cost Rates fourth place in economy out of five processes
SOLVENT EXTRACTION	Sulfur Light oil Coal extract Carbon dioxide Fuel gas	Requires no pyrite concentration; uses entire refuse stream Produces high quality products	Complex process; it requires further development and solution of technical problems Economics favor very large-scale operation	Adds $1.63/ton to clean coal cost Least economical process
FLUID-BED ROASTING	Sulfuric acid Iron oxides Electric power	Relatively simple, well-developed technology	Requires pyrite concentration Requires pilot plant development for use with coal refuse Dilution of pyrite with coal causes lower concentration of SO_2 in gas stream. Degree of acid discoloration is unknown	Adds $0.12/ton to clean coal cost Rates second place in economy out of five processes
OUTOKUMPU PROCESS	Sulfur Sulfuric acid Iron oxides	Recovers about half of the sulfur in elemental form	Ash content of coal refuse would cause technical problems in furnace Economy would suffer because iron sinter would not be marketable Requires very large-scale operation	Adds $0.41/ton to clean coal cost Rates third place in economy out of five processes

Low-temperature carbonization-desulfurization of coal refuse appears to be more expensive than the three processes already described. The high capital cost and doubtful market for the resulting high-ash char result in a negative outlook for this process.

The high capital cost of the solvent extraction process makes it the most expensive of the processes. The high-ash content of the pyrite-coal feed material is partially responsible for the high capital cost. The controlling factor in the capital cost, however, is the large number of individual process operations which are necessary in a complete solvent extraction plant. Determination of the cost of this process at much higher feedrates indicates that the throughput must be increased by a factor of 3 to 5 before it shows promise. The costs currently available are highly uncertain and should be evaluated when Pittsburgh-Midway complete their pilot plant program.

Table 6-1 shows that the cost of transporting the by-products to market, especially sulfuric acid, is significant in the economics. This affects the selection of a process and the plant location.

The relatively promising outlook for direct firing of refuse is related to the following factors:

- The heating value of the refuse is recovered efficiently in a simple manner.
- The SO_2 concentration of the combustion gases is up to ten times as high as the concentration in typical power plant stack gases. Increased product value outweighs the increase in recovery cost.

Tables 6-3 through 6-7 give the yearly operating cost breakdown for the processes compared in Table 6-1.

Table 6-3

OPERATING COST SUMMARY

COMBUSTION-SULFUR OXIDES REMOVAL
(MONSANTO PROCESS)

Basis	
Refuse coal feed (dry basis), short T/yr	630,000
Refuse coal composition (dry basis), % coal	67.4
% ash	32.6
% sulfur	6.2
Sulfuric acid (77.7% H_2SO_4), short T/yr	116,000
Power plant rating, Mw	250
Power plant and process plant annual load factor, %	65
Power plant heat rate, Btu/kwhr	9,000
Annual power generation, millions of kwhr	1,425
Process plant capital cost, $	8,300,000
Operating Cost	**$/yr**
Power (1190 kw x 5700 hr x $0.008)	54,000
Catalyst (150,000 liters at $1.40/liter, 2-year life)	105,000
Supplies (0.15%/yr of capital cost)	13,000
Chemicals and laboratory	3,000
Waste disposal (218,000 T/yr ash at $0.75/T)	163,000
Operating labor (48 manhours/day at $4.00/hr x 365)	70,000
Supervision and overhead	50,000
Maintenance (6% yr on $8,300,000)	498,000
Capital charges (15% yr on $8,300,000)	1,245,000
Total Yearly Cost	2,201,000
Credits	
Sulfuric acid (77.7%) 116,000 T/yr at $10/T	1,160,000
Fuel value: 630,000 T/yr at $1.95/T	1,230,000
Total Credits	2,390,000
Net Yearly Cost	-189,000

Change in Clean Coal Price, $/T: -0.10

Table 6-4

OPERATING COST SUMMARY

CARBONIZATION-DESULFURIZATION
(CONSOL PROCESS)

Basis	
Refuse coal feed (dry basis), short T/yr	630,000
Refuse coal composition (dry basis), % coal	67.4
% ash	32.6
% sulfur	6.2
Plant operating days/yr	330
Elemental sulfur, long T/yr	29,000
Char (59% coal, 1% sulfur, 38% ash), short T/yr	445,000
Pipeline gas (976 Btu/Scf), Scf/hr	433,000
Process plant capital cost, $	14,000,000
Operating Cost	**$/yr**
Power (640 kw x 7900 hr x $0.008)	41,000
Cooling water (15,000 gpm at $0.05/M gal)	355,000
Boiler feed water (15 gpm at $0.50/M gal)	4,000
Labor (40 men at $4.00/hr)	333,000
Supervision and overhead	333,000
Maintenance (6%/yr on $14,000,000)	840,000
Capital charges (15%/yr on $14,000,000)	2,100,000
Total Yearly Cost	4,006,000
Credits	
Elemental sulfur: 29,000 long T/yr at $36.30/T	1,050,000
Char: 445,000 short T/yr at $1.10/T	490,000
Pipeline gas: 7,220 Scfm (976 Btu/cf) at $0.26/Mscf	892,000
Total Credits	2,432,000
Net Yearly Cost	1,574,000

Change in Clean Coal Price, $/T: 0.84

Table 6-5

OPERATING COST SUMMARY
SOLVENT EXTRACTION PROCESS

Basis	
Refuse coal feed (dry basis), short T/yr	**630,000**
Refuse coal composition (dry basis), % coal	**67.4**
% ash	**32.6**
% sulfur	**6.2**
Plant operating days/yr	**330**
Elemental sulfur: long T/yr	**18,700**
Light oil, gal/yr	**5,700,000**
Process gas, Mscf/yr	**1,490,000**
Process plant capital cost, $	**19,500,000**
Operating Cost	$/yr
Power (4720 kw x 7900 hr x $0.008)	300,000
Cooling water (10,000 gpm at $0.05/M gal)	240,000
Fuel (130,000 T/yr of ROM coal at $3.85/T)	500,000
Ash disposal (210,000 T/yr at $0.75/T)	160,000
Labor (128 men at $4.00 hr)	1,065,000
Supervision and overhead	1,065,000
Maintenance (6%/yr on $19,500,000)	1,170,000
Capital charges (15%/yr on $19,500,000)	2,920,000
Total Yearly Cost	7,420,000
Credits	
Light oil: (5,700,000 gal/yr at $0.10/gal)	570,000
Process gas: (3140 Scfm at $0.26/Mscf)	388,000
Sulfur: (17,000 long T/yr at $36.30/T)	620,000
Carbon dioxide: (52,000 T/yr at $15/T)	780,000
Coal extract: (315,000 T/yr at $6.40/T)	2,010,000
Total Credits	4,368,000
Net Yearly Cost	3,052,000

Change in Clean Coal Price, $/T: 1.63

Table 6-6

OPERATING COST SUMMARY
FLUID-BED ROASTING

Basis	
Feed material, 90% pyrite concentrate, T/yr	36,100
Acid (98%) production, T/yr	51,600
Percentage of sulfur recovery	95
Plant operating days/yr	330
Fluidized roasters with waste heat boilers	
Plant investment, $	2,400,000
Operating Cost	
Electricity (105 kwhr/T acid at $0.008/kwh)	43,000
Boiler water (400 gal/T acid at $0.50/M gal)	25,000
Cooling water (3080 gal/T acid at $0.05/M gal)	8,000
Lime	5,000
Waste disposal (24,000 T/yr calcines at $0.75/T)	18,000
Labor (22 men at $4.00/hr)	183,000
Supervision and overhead	183,000
Maintenance (6%/yr on $2,400,000)	144,000
Capital charges (15%/yr on $2,400,000)	360,000
	969,000
Cost of Concentrating Feed (36,100 T/yr at $3.00/T)	108,000
Total Yearly Cost	1,077,000
Credits	
Sulfuric acid (98%): 51,600 T/yr at $16/T	826,000
Steam Value: 9000 lb/hr at $0.50/M lb	36,000
Total Credits	862,000
Net Yearly Cost	215,000

Change in Clean Coal Price, $/T: 0.12

Table 6-7

OPERATING COST SUMMARY
OUTOKUMPU PROCESS

Basis	
Feed material, 90% pyrite concentrate, T/yr	36,100
Acid (98%) production, T/yr	19,000
Elemental sulfur production, long T/yr	7,400
Iron oxide pellets, T/yr	22,600
Plant operating days/yr	330
Plant investment, $	5,000,000
Operating Cost	**$/yr**
Fuel (6000 T/yr clean coal x $5.50/T)	33,000
Cooling water (2100 gpm at $0.05/M gal)	50,000
Catalysts and chemicals	38,000
Labor (25 men at $4.00/hr)	208,000
Supervision and overhead	208,000
Maintenance (6%/yr on $5,000,000)	300,000
Capital charges (15%/yr on $5,000,000)	750,000
	1,587,000
Cost of Concentrating Feed (36,100 T/yr at $3.00/T)	108,000
Total Yearly Cost	1,695,000
Credits	
Sulfuric acid (98%): 19,000 T/yr at $16/T	304,000
Sulfur: 7400 long T/yr at $36.30/T	269,000
Iron oxide pellets: 22,600 T/yr at $13/T	294,000
Steam value: 15,250 lb/hr at $0.50/M lb	60,000
Total Credits	927,000
Net Yearly Cost	768,000
Change in Clean Coal Price, $/T: 0.41	

6.2 BASIS AND CRITERIA OF ESTIMATES

The preliminary economic evaluation of all the foregoing processes was based on the following series of criteria. The practices in the preliminary estimates were continued into the detailed economic analysis of the selected process. It will be obvious from the criteria and from the nature of their use that the analysis must be considered no better than a rough order of magnitude, and that factors not immediately important may conceivably impact the estimates significantly.

- Coal Mine: The coal mine produces 2,500,000 tons per year of ROM coal. Mining cost is $3.50 per ton.
- Coal Preparation Plant: The coal preparation plant produces 1,870,000 tons per year of clean coal (1.0% sulfur) at $0.50 per ton. This quantity is enough to fuel a 1000 Mw(e) electric power generator.
- Clean Coal: The clean coal has a heating value of 13,750 Btu per pound, a total sulfur content of 1 percent, and an ash content of 4 percent.
- Pyrite-Coal Refuse: A 630,000-ton-per-year stream of pyrite-coal refuse is produced containing 33 percent mineral matter and 67 percent coal. The total pyritic and organic sulfur content is 6.2 percent, and the heating value is 9700 Btu per pound.
- ROM Coal: The ROM coal has a heating value of 13,000 Btu per pound and contains 0.8 percent organic sulfur, 1.5 percent pyritic sulfur, and 10 percent ash.
- Pyrite Concentration: For the processes which require it, a pyrite concentration plant processes 1900 tons per day of pyrite-coal refuse, and produces 110 tons per day of pyrite containing 90 percent FeS_2, 5 percent coal, and 5 percent mineral matter with a pyrite recovery efficiency of 50 percent and a total operating cost of $3.00 per ton of concentrate.
- Capital Costs: Capital costs for the pyrite-coal utilization plant are the installed battery limit costs of the process equipment and auxiliary facilities. The cost of land is not included in the estimate. The installed costs are derived from estimates published in nonproprietary sources using published scale factors to adjust for plant size.

- Utilities: The unit cost of utilities required for operating these processes are:

Cooling Water	\$0.05/1000 gallons
Boiler Feedwater	0.50/1000 gallons
Electric Power	0.008/kwhr

- Raw Materials: The unit cost of raw materials required for operating these processes are:

ROM Coal	\$3.85/ton-13,000 Btu/lb 3.50/ton mining cost
Clean Coal	5.50/ton-13,750 Btu/lb 75% recovery
Pyrite Concentrate	3.00/ton-90% FeS_2
Ash Disposal Cost	0.75/ton

- Labor: Operating labor rates are averaged at \$4.00/hr.

- Supervision and Overhead: The prorated cost of supervision, employee benefits, and administrative overhead is equal to operating labor cost.

- Maintenance, Labor, and Material: An amount equal to six percent of plant capital cost is assumed for annual maintenance, labor, and material costs.

- Fixed Capital Charges: The annual fixed charge rate is 15 percent of the pyrite-coal utilization plant capital cost. This includes 20-year depreciation at 6.5 percent interest, property taxes, insurance, and administrative expenses.

- Transportation Costs: The cost of shipping products to a market 200 miles away, based on prevailing railroad point-to-point commodity shipping rates for the various products, is as follows:

Clean Coal (unit train)	\$ 1.44/ton
Coal Extract	2.90/ton
Coal Char	2.90/ton
Iron Oxide	6.25/ton
Sulfur	8.25/ton
Sulfuric Acid	11.00/ton
Sulfur Dioxide	12.00/ton
Light Oil	8.00/ton
Carbon Dioxide	8.00/ton

- By-product Credits: Credits for by-product sale, FOB plant, are derived from the prevailing market selling prices for each by-product. The credits are:

Sulfuric Acid (98%)	$16.00/ton	From 500 T/day plant using elemental sulfur at $38.00/ton producing 98% acid
Sulfuric Acid (77.7%)	10.00/ton	From 98% acid at $16.00/ton and discount 20% for quality
Elemental Sulfur	36.30/long ton	Liquid crude, dark sulfur, midwest plant
Coal Extract	6.40/ton	15,800 Btu/lb based on clean coal at $5.50/ton
Char	1.10/ton	8600 Btu/lb based on ROM coat at $3.50/ton and discounted 50% for quality
Fuel Value of Coal Refuse	1.95/ton	9700 Btu/lb
Pipeline Gas	0.26/1000 scf	975 Btu/scf
Light Oil	0.10/gal	Pittsburgh-Midway
Iron Oxide Pellets	13.00/ton	67% Fe
Liquid Carbon Dioxide	15.00/ton	Pittsburgh-Midway and Air Reduction Co.
Steam	0.50/1000 lb	

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Section 7

REFUSE COMBUSTION - SULFUR OXIDES RECOVERY

Refuse combustion-sulfur oxides recovery uses coal refuse by direct firing in a special steam generator to produce electricity; sulfur oxides are then converted to sulfuric acid by the Monsanto Catalytic Oxidation process. The preliminary analysis of Section 6 showed that this scheme for refuse processing had the most promising economics of those under consideration. From a technical viewpoint, the method appears to be relatively simple and based on sound development work. For these reasons, it was selected for further analysis. In the following section, this analysis is presented under two major headings: Technical Aspects and Economics.

7.1 TECHNICAL ASPECTS

7.1.1 Process Description

Figure 7-1 presents a schematic concept of the Monsanto process for removing sulfur oxides from flue gas. A description of the major steps follows.

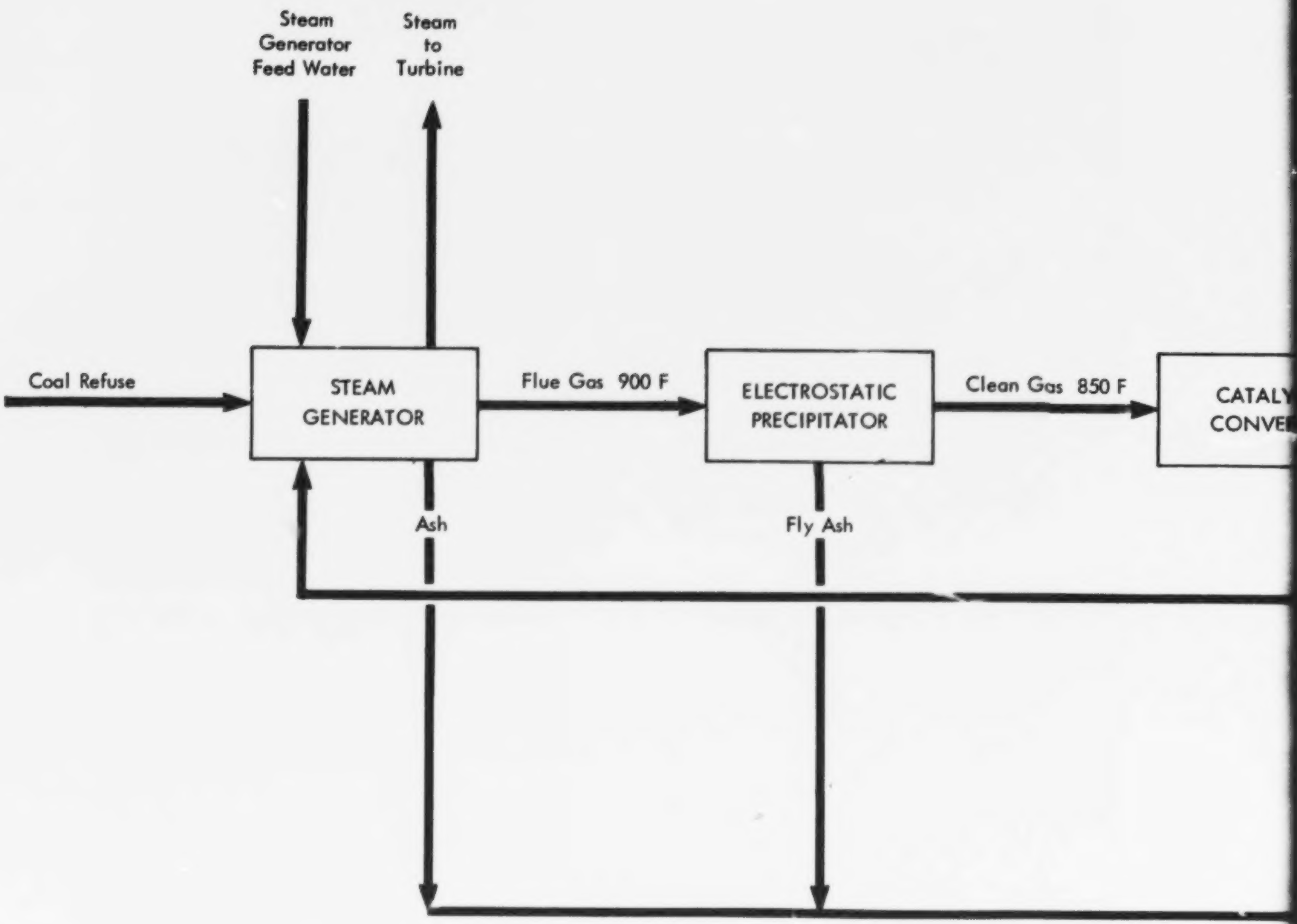
Gas Cleaning. Particulates in the flue gas will plug or poison the catalyst, thereby reducing its activity and its life. Maximum tolerable dust loading of the gas is about 0.003 to 0.006 grain per scf. For a coal-fired boiler, this cleaning requirement implies a collection efficiency above 99.5 percent. Pilot plants burning oil have operated with only an electrostatic precipitator. Plants capable of coal firing have used a mechanical collector ahead of an electrostatic unit. The high ash content of coal refuse would probably require the two-step collection system.

Catalytic Oxidation. To obtain optimum conversion of sulfur dioxide to trioxide, the temperature of the gas and catalyst should be about 800-850 F. This implies that flue gas leaves the boiler at 850-900 F and drops about 50 F through the dust-collection system. Vanadium pentoxide catalyst promotes the reaction between the sulfur dioxide and oxygen in the flue gas at an optimum conversion efficiency of about 90 percent.

Pilot plant work appears to favor a fixed bed of catalyst pellets; one half the catalyst is replaced per year. The catalyst performance and life comprise one of the sensitive factors in the Monsanto process.

Condensation. Gas from the catalytic converter is cooled from 850 to 250 F in a rotary air preheater. As the temperature falls, SO_3 reacts with water vapor in the flue gas, forming sulfuric acid. The reaction proceeds rapidly to completion below 400 F. Acid vapor condenses on the corrosion-resistant surface of the air preheater and is collected. Some condensed acid is entrained in the gas stream and removed in the next step.

An alternative method of acid recovery is to use a wet scrubbing system to condense acid rather than the metal heat-exchanger surface.


Steam
Generator
Feed Water
Steam
to
Turbine
Coal Refuse
STEAM
GENERATOR
Flue Gas 900 F
ELECTROSTATIC
PRECIPITATOR
Clean Gas 850 F
CATALY
CONVE
Ash
Fly Ash

A

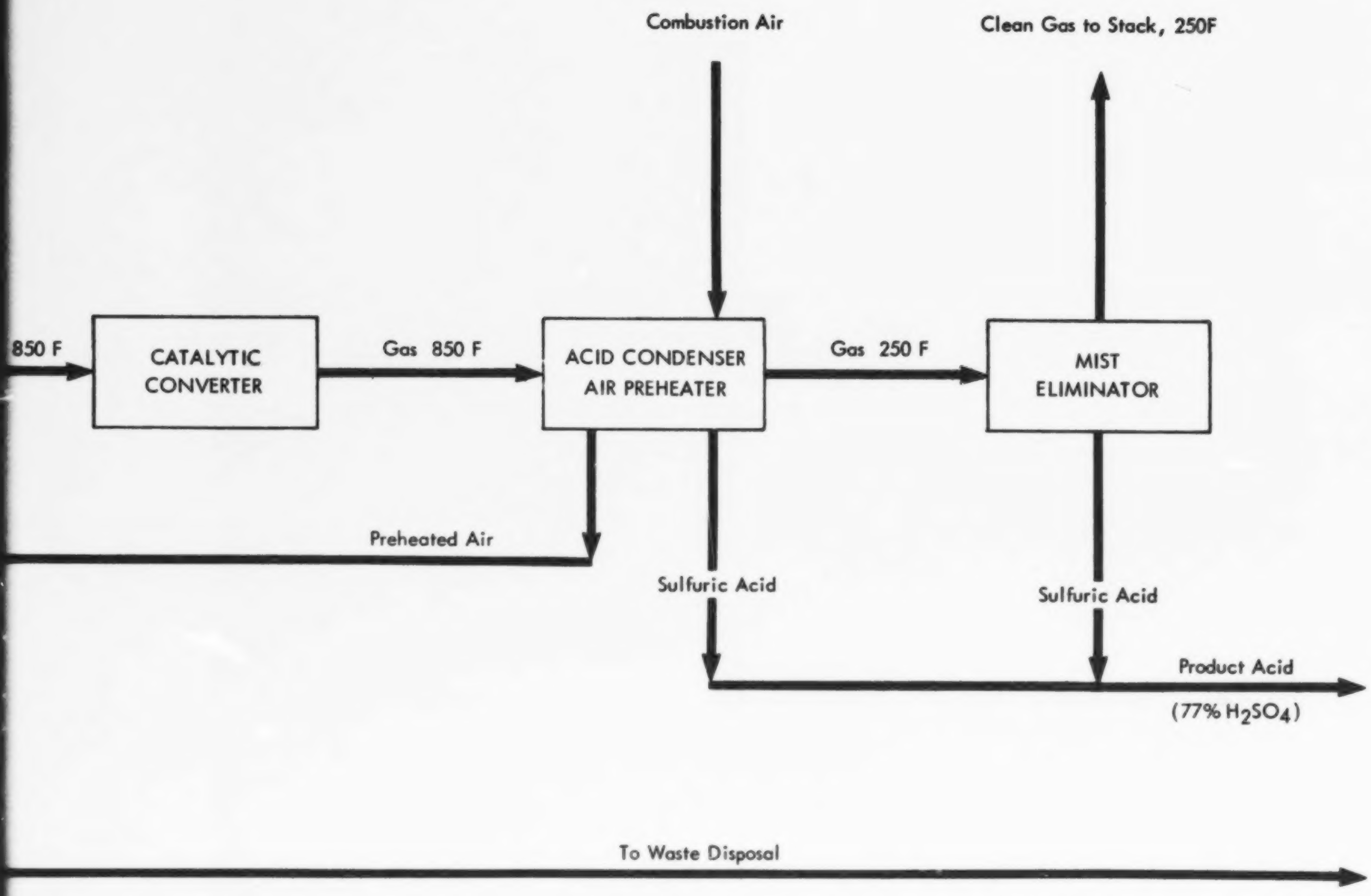

Combustion Air
Clean Gas to Stack, 250F
850 F
CATALYTIC CONVERTER
Gas 850 F
ACID CONDENSER AIR PREHEATER
Gas 250 F
MIST ELIMINATOR
Preheated Air
Sulfuric Acid
Sulfuric Acid
Product Acid
(77% H2SO4)
To Waste Disposal

Figure 7-1. Combustion-Sulfur Oxides Removal

B

In the wet system, the air preheater would cool the gases from the converter to a temperature slightly above the dew point. The gas would then pass into an absorber, operating with circulating acid. Condensation of the acid content of the gas stream would produce 78-80 weight percent H_2SO_4, about the same strength as from the dry system.

A major operating problem associated with acid condensation is corrosion. Materials in contact with the gas stream at temperatures below the dew point of the acid are subject to severe corrosive attack. Use of materials to minimize corrosion, such as Cor-Ten, would significantly increase capital costs. The trend in the process development seems to favor ceramic or enamel coatings or plastics rather than corrosion-resistant metals.

Mist Elimination. Gas leaving the preheater-condenser contains entrained, micron-sized, sulfuric acid mist, which must be reduced below 30 ppm to avoid discharging a visible plume to the atmosphere. Monsanto reports excellent results with a Brink type mist eliminator. A typical element of this eliminator consists of a packed bed of coated glass fibers held in a mounting frame by wire or plastic screen. The sulfuric acid collected from this step is combined with that from the condenser.

Under cold-weather conditions, the temperature of the flue gas exhausting to the plant stack should not fall much below 250 F to avoid visible steam plume.

Product. Sulfuric acid collected from the mist eliminator and the preheater-condenser gives a product concentration of about 60 degrees B'e (77 weight percent). The acid is equal to commercial grades but with slight discoloration.

Pressure Drop. The estimated flue-gas pressure drop through the equipment used for the Monsanto process, not including the boiler, is 20 to 30 inches of water. This pressure drop must be made up by a fan located somewhere between the mist eliminator and the stack.

Boiler Modifications. The typical temperature of flue gas leaving the economizer of a modern boiler is about 765 F. Because of the catalytic conversion requirements, the gas should leave the boiler at 850-900 F. This would require a change in operation or location of the typical economizer: either the economizer can be bypassed or gas exiting the catalytic converter can flow through the economizer.

Sulfur Oxides Removal Efficiency. The overall sulfur oxides removal efficiency of the Monsanto process is about 85 percent. This requires that individual steps of the process have high efficiencies. For example, if the efficiency of catalytic conversion is 90 percent, the combined efficiency of acid formation and removal must be at least 95 percent.

7.1.2 Steam Generator System

Firing a solid fuel with such high ash and sulfur content as coal refuse requires special consideration in designing the steam generator. It was pointed out before that the Monsanto process requires a boiler modification concerning the position and use of the economizer However, ash characteristics and quantity impose the most important design criteria. More specifically, the problem concerns the method of removing ash from the furnace. The discussion which follows describes ash properties and their relationship to boiler design. It further attempts to emphasize the significance of this technology to firing coal refuse.

Ash Properties. Ash consists of inorganic compounds which (except for pyrite, FeS_2) are not combustible. This mineral matter must be carried through the coal burning process and ultimately removed for disposal. Since coal and its contained ash are so widely dispersed over the earth, the relative amounts of mineral components vary widely. Ash occurs chiefly as slate, clay, sandstone, limestone, and pyrite. The quantitative evaluation of the exact mineral species is extremely difficult. In practice, a simplified ASTM method gives an analysis in terms of oxides of elements (SiO_2, Al_2O_3, Fe_2O_3, CaO, MgO, Na_2O, and K_2O).

Ash fusion temperature is of special interest in steam-generator design since it indicates whether or not the ash can be removed as a molten slag. Ash does not have a sharply defined melting point like ice or most metals. Its fusion characteristics are more like those of glass and extend over a range of temperatures. Actually, three temperatures are determined in the laboratory, each having a special significance to the design and operation of specific areas within the boiler. Definitions of these temperatures follow:

- Initial Deformation Temperature (IDT) - at this temperature, particles of ash have a slight tendency to stick together or to build up slowly on heat-absorbing surfaces.
- Softening Temperature (ST) - at this temperature ash particles show a greatly accelerated tendency to mass together and stick in large quantities to heat-absorbing surfaces.
- Fluid Temperature (FT) - at this temperature, coal ash will flow in a stream or drip from heat-absorbing surfaces.

The presence of iron in coal ash affects its fusion in an interesting way. Completely oxidized iron (ferric oxide, Fe_2O_3), raises all three values of ash fusion temperature, while the ferrous form (FeO)

lowers them. The standardized procedure for determining fusion temperatures calls for a reducing atmosphere in the testing furnace.

Although fusion temperatures are convenient indications of ash behavior, viscosity constitutes a more dependable guide for evaluating removal of ash in liquid form from the furnace. Actual measurement of viscosity requires considerable time and equipment. Such data are not commonly available. However, there has been some success in correlating silica content with viscosity of coal-ash slags. The temperature of critical viscosity can be determined experimentally and is an important property for evaluating a slag. This defines the transition point between liquid and plastic stages and sets the lower temperature limit for design considerations. Calcium compounds and ferrous iron are the principal fluxing components of the ash and can be used as parameters in viscosity-temperature correlations.

The relationship between coal-ash and boiler design will be discussed from the standpoints of three methods of fuel firing that could apply to coal refuse. These methods are: pulverized coal firing, cyclone firing, and fluid-bed combustion.

Pulverized Coal Firing. Pulverized coal is now the most widely used fuel for electric power generation. Its success is due to the development of efficient methods of firing and of very flexible boiler design. Almost any coal can be burned successfully in pulverized form. Many low-grade and waste solid fuels are efficiently fired in this manner.

The degree of coal fineness is one of the key factors in firing pulverized coal. This requirements varies from coal to coal. In general, highly volatile coals which ignite more readily do not require as fine pulverization as less volatile ones. Degree of fineness ranges from

65 to 85 percent minus 200 mesh. The ease with which a coal can be pulverized can be determined by an ASTM test.

On the basis of ash properties, furnaces for pulverized coal firing are specifically designed for dry or liquid ash removal. This problem becomes one of paramount interest in firing coal refuse with its extremely high content of ash and sulfur. Currently, the dry ash method predominates because it can adapt to a wider variety of coals having considerable range in ash fusion temperatures.

Ash fusion and viscosity characteristics affect this phase of furnace design more than any other factors. Coal ash having an ST above 2600 F or an ash viscosity greater than 250 poises at 2600 F are suited to dry-ash removal; ashes with an ST below 2200 F are suited to removal as liquid slag. Most coals have intermediate characteristics but can be used with either method if the boiler has means of correctly controlling temperatures within its different zones to minimize slagging on furnace walls and tubing.

For slag-tap furnaces, a smaller combustion zone is used than for dry ash operation, the burners concentrating heat close to the floor and tap hole. Such a furnace must be able to withstand temperatures over 3000 F. Tapping of viscous slags can be aided by adding fluxes (compounds of calcium, iron, or alkali metals) to the fuel or slag pool. Slags having considerable iron can be made more fluid by decreasing excess air to retain the ferrous state of oxidation. However, such conditions could cause incomplete oxidation of pyrite and result in accumulation of molten iron sulfide. This could damage floor structure and decrease sulfur recovery.

When pulverized coal burns in a dry ash furnace, the flue gases may entrain and carry over 85 percent of the total ash in the coal. A slag-tap furnace may retain as much as 50 percent of the ash as fluid slag which facilitates disposal and decreases the size and cost of dust-collecting equipment.

Cyclone Furnace. This is a special slag-tap furnace particularly adaptable to high-capacity units and to use of low-grade coals. Cyclone firing was developed more recently than pulverized coal firing, but has received wide acceptance in the United States and Germany. However, its application is limited to certain ash characteristics.

The cyclone furnace is a water-cooled cylinder, set to slope a few degrees from horizontal, into which crushed fuel (1/4-inch or less) and combustion air enter tangentially. Ash melts and forms a thin film which moves along the furnace wall. Incoming fuel particles (except the fines that burned in suspension) are thrown to the wall and caught in the running slag film. Secondary air sweeps past these embedded particles at high speed, causing rapid burning of coal and melting of ash. Molten slag discharges through a tap hole.

A cyclone furnace appears to be a particularly appropriate design for refuse firing if ash characteristics are suitable. Its range in the use of fuels is broad, far exceeding that of slag-tap pulverized coal furnaces. A fuel should be satisfactory for firing if the volatile matter exceeds 15 percent (dry basis), the viscosity of the slag is less than 250 poises at 2600 F, and the ash softening temperature is not higher than 2500 F. An increase in the temperature of the combustion can extend the limits of the coal characteristics. Generally, the poorer grades of coals are the best suited for cyclone-furnace firing.

Some of the important features and advantages of cyclone firing are summarized here:

- Fine grinding is not necessary; this reduces the required investment below that for pulverized coal firing.

- Most of the ash is retained as liquid slag in the furnace bottom; this reduces deposition in boiler convection passes or emission as fly ash. The quantity of fly-ash emission is only 10 to 15 percent of the total ash in the coal. (Pulverized coal firing has 40 to 60 percent carryover with slag-tap, and 80 to 90 percent with dry firing.)
- About 85 percent of the fly-ash from the cyclone furnace is smaller than 10 microns. These small particles settle slowly and disperse widely, which tends to reduce ground level concentrations.
- Drastic reduction in fly-ash decreases the size and cost of dust-collection equipment.
- Tapped slag can be disposed of much easier than collected dust. In some geographical locations, this material may have some value as fill or light-weight aggregate.
- Combustion-furnace volume requirement is small.
- Because of low dust load and small particle size, high gas velocities can be used in the convection heat-absorbing zones. This increases efficiency of heat transfer without danger of tube erosion.

Fluid Bed Combustion. Conventional coal-fired power stations, which use pulverized coal or cyclone firing, have reached a very high degree of efficiency. Consequently, there is little chance for significant improvement in design and economics for such plants. As a result, a new firing technique, fluid-bed combustion, is under development in an effort to make coal meet the competition from other fuels.

This method is under development by at least two groups of researchers. One group consists of the American consulting engineering firm of Pope, Evans, and Robbins under the sponsorship of the Office of Coal Research. Their goal is to develop a factory-assembled, or packaged, coal-fired boiler that would compete with oil or gas-fired units.

Another research group, under the National Coal Board of England, has also developed a fluidized-bed combustor. Experimental results from both groups have been encouraging, and this technique is of definite interest for firing refuse.

Figure 7-2 shows one concept of a fluid-bed steam generator. Coal particles (3/4-inch or less) are fed pneumatically into the bottom of a fluidized bed of coal ash. Air for combustion and fluidization is forced into the windbox of the furnace, then up through a distribution grid into the bed. At start-off, the bed is heated to ignition temperature by external means. Thereafter, the temperature is maintained by adjusting combustion conditions and steam rate.

About two-thirds of the recoverable heat from combustion is transmitted into the water-tube envelope around the furnace walls by direct contact. Heat transfer from a turbulent bed to a heating surface is 5 to 10 times as fast from gas to surface. Hot gases then

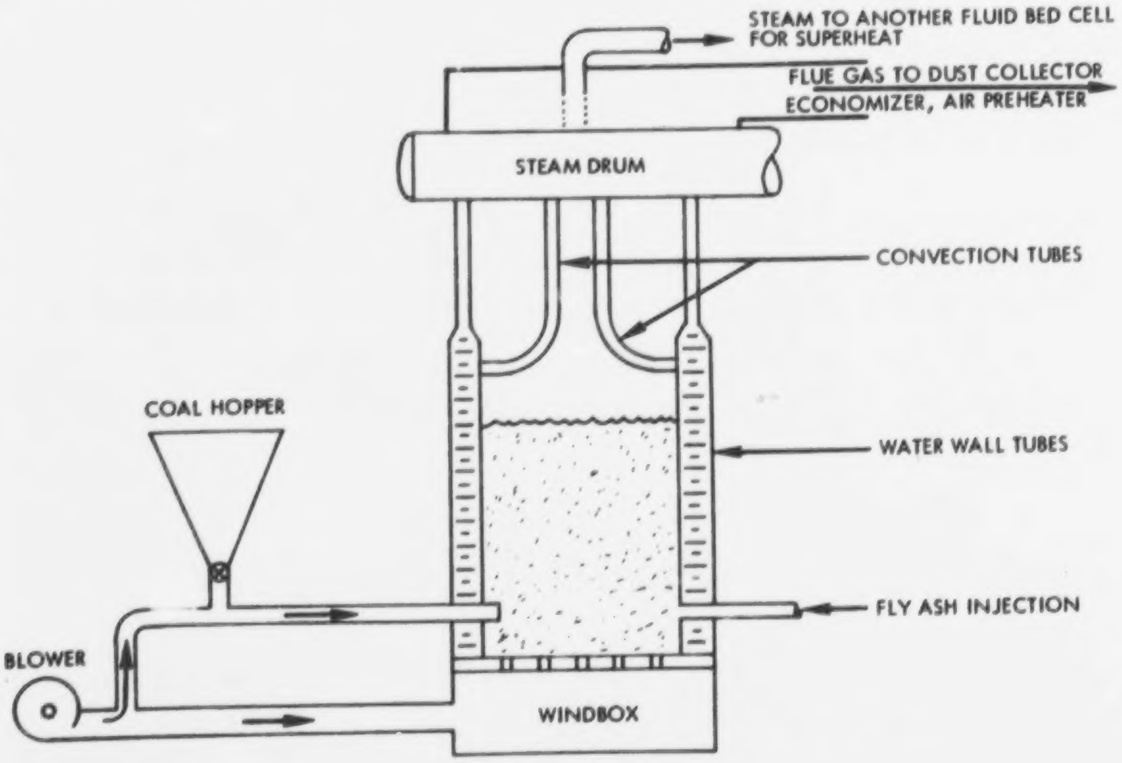

STEAM TO ANOTHER FLUID BED CELL FOR SUPERHEAT
FLUE GAS TO DUST COLLECTOR ECONOMIZER, AIR PREHEATER
STEAM DRUM
CONVECTION TUBES
COAL HOPPER
WATER WALL TUBES
FLY ASH INJECTION
BLOWER
WINDBOX

Figure 7-2. Fluid-Bed Steam Generator

flow past convection boiler tubes, dust collectors, economizer, and preheater. Fly ash returns to the furnace by air injection. A superheater surface can be immersed directly in the oxidizing fluid bed. A series of oxidizing fluid-bed modules can be assembled to provide whatever size boiler unit is required.

The lower limit for keeping the bed ignited ranges from 1200 F for highly volatile bituminous coal to 1600 F for anthracite. The upper limit depends upon ash fusion temperatures. Generally, the bed cannot exceed about 150-200 F below IDT without becoming sticky. Pilot plants were able to sustain beds over 2000 F with high-fusion coals but this decreased to 1900 F for certain low-fusion, high-sulfur coals.

It must be emphasized that fluid-bed combustion with steam generation is still in development stages. From a theoretical standpoint, this method promises some important advantages over conventional fuel firing, and it may prove particularly suitable for firing refuse. Some of the advantages are summarized here:

- Capital cost will be lower because of the smaller, more compact unit. The high heat transfer coefficient should result in fewer tubes. Highly efficient fluid-bed combustion results in high heat release values and very low combustion volume requirements. The overall effect may be to reduce the size of the boiler to 1/4 that of a conventional type.
- The cost of grinding fuel will be lower. Fluid beds can operate on much coarser feed than a pulverized coal boiler.
- It is possible to attain much better oxygen utilization; excess air can be 2-5 percent. This results in higher sulfur oxides concentration in the flue gas.

- Tube erosion has been nil in the pilot plant; this is credited to low gas velocity (3-4 fps).
- Because of relatively low combustion temperatures, the volatization of alkali salts from the ash will be small, thus decreasing corrosion from condensation on cooler surfaces.

Some of the current problems which exist and are receiving attention in the pilot plants are listed below:

- Initial work resulted in carbon losses in fly ash amounting to 20 percent of the heat input. This has been reduced to 5 percent. Efforts continue for further reduction.
- Finer particle sizes give better heat transfer but result in more elutriation to the dust collectors. It will be necessary to arrive at an economic balance among the factors: heat transfer surface, grinding cost, quantity of ash, cost of dust collectors, effect of dust on catalyst life.

Some preliminary estimates of capital and operating costs for fluid-bed steam generation should be developed for large-scale operation. The expense of recirculating flue gas for fluidization may prove to be a major component of operating cost.

7.1.3 Other Processes for Sulfur Oxide Removal

Numerous processes are under development for removing sulfur oxides from flue gases. The Monsanto process was picked for the economic evaluation, largely on the basis of its sound, relatively simple technology. Comparison of preliminary economics showed that it compares favorably with other processes, especially when applied to a high-sulfur fuel.

A detailed economic analysis of competing SO_2 removal processes is outside the scope of this report. Such an effort specifically applied to coal refuse could lead to a substantial improvement in the overall economic outlook. As a first step in this direction, four processes that may compete with Monsanto will be briefly discussed.

Certain classes of processes are not considered suitable for use with coal refuse combustion.

- Processes that produce a byproduct such as ammonium sulfate, which has limited market appeal.
- Processes that produce no salable byproduct, such as the alkali injection processes. With coal refuse burning, a byproduct credit is the only hope of making the effort of sulfur recovery attractive.

Very few conclusions have been published regarding the economic sensitivity of these processes to fuel sulfur content. Information that has been developed applies to fuel sulfur contents ranging from 1.5 to 4.5 percent rather than the much higher values characteristic of coal refuse.

Topsoe Process. This process is used in the Lacq natural gas field in France to recover additional sulfur value from gases exhausted from the Claus sulfur recovery plants. As used in France, the basic process steps are:

- Incineration - Gases from the Claus plants are burned in air to convert all sulfur compounds to sulfur dioxide
- Catalytic Conversion - Sulfur dioxide is converted to trioxide as in the Monsanto process

- Absorption - The SO_3 rich gases are cooled to a temperature slightly above their dew point, then absorbed in dilute recycle acid.
- Acid Concentration - Heat recovered from the catalytic converter gases is used to concentrate the dilute acid (70-80 percent) to 92-94 percent.

Part of this process could be used as a modification to the Monsanto scheme. The wet absorption step of the Topsoe process could replace condensation by air preheater in the Monsanto process. Use of waste heat to concentrate acid would have to be weighed against loss of some preheat in the combustion air.

Alkalized Alumina. This process uses alkalized alumina pellets to absorb sulfur dioxide from flue gas. The pellets are regenerated and elemental sulfur is recovered.

The process was initially developed by the U.S. Bureau of Mines under sponsorship of the U.S. Public Health Service. Other work on this process was done by the British Central Electricity Generation Board. Shell Chemical Company has a foreign patent on a similar process with a variation on the absorbent. Basic steps of the process follows:

- Particulate Removal - Flue gas from the boiler must pass through high-efficiency dust collectors to minimize contamination of the absorbent.
- Absorption - Sulfur oxides are absorbed from the flue gas on alkalized alumina pellets at 525 F in a transport or fluid-bed reactor. The Bureau of Mines absorbent is a co-precipitate of sodium and aluminum oxides (about 20-percent sodium oxide).
- Absorbent Regeneration - The spent absorbent is regenerated at about 1000 F with a reducing gas such as hydrogen and carbon monoxide. Reactions

form hydrogen sulfide, carbonyl sulfide, and water vapor.

- Sulfur Recovery - Hydrogen sulfide and carbonyl sulfide yield elemental sulfur in a Claus unit.

Preliminary estimates show that the capital cost of the Alkalized Alumina process is about half that of the Monsanto process, but the operating cost is about the same. The absorbent material is quite expensive ($500/ton) so the process is sensitive to make-up requirements. Some published economics have shown that net operating costs increase with sulfur content of coal except for very high sales value of sulfur.

Reinluft Process. This process uses char to remove sulfur dioxide from flue gas. The sulfur dioxide is recovered and converted to sulfuric acid. Large-scale Rienluft pilot plants have operated in Germany and Japan. Basic process steps are:

- Adsorption - Flue gas enteres the bottom of a bed of activated char, where SO_3 is removed. The flue gas is then withdrawn, cooled to 220 F, and returned to the adsorber at a higher level. SO_2 in the gas oxidizes to SO_3 and reacts with adsorbed water in the char to form sulfuric acid.
- Desorption - Char with its adsorbed acid drops to the regenerator where its temperature is raised to 700 F by recycle inert gas. The sulfuric acid dissociates into SO_3 and H_2O, and the SO_3 reacts with carbon to form CO_2 and SO_2. A product stream containing about 25 percent SO_2 is withdrawn from the recycle gas and sent to acid production. Regenerated char is screened to remove dust and is recycled.
- Acid Production - Concentrated sulfuric acid can be produced by a conventional contact plant.

A difficult operating problem concerns uncontrolled oxidation of carbon by stack gas oxygen and development of hot spots. Other problems include the expense of recirculating large amounts of carbon and costs of heat and the reducing agent.

Preliminary estimates show that the capital cost of the Reinluft process is about the same as the Monsanto process, but the operating cost is higher.

Wellman-Lord Process. This process uses a regenerative chemical absorbent to recover anhydrous sulfur dioxide from flue gas. A pilot unit was installed on a commercial power station and operated for five months. A larger unit is now in the planning stage. Basic steps of the process follow.

- First Stage Gas Cleaning - Most of the fly ash in the flue gas is removed in an electrostatic precipitator.
- Second Stage Gas Cleaning - SO_3 and the remaining fly ash are removed by wet scrubbing and discarded as a low-volume effluent.
- SO_2 Absorption - SO_2 is absorbed from the dust free gas by an alkali solution. SO_2 free gas is now saturated with water vapor at about 120 F. Under certain weather conditions, the gas may have to be reheated to avoid a visible steam plume at the stack discharge.
- Solution Regeneration - SO_2 is steam stripped from the solution and condensed in anhydrous form. The regenerated solution recycles to the absorption step.

Although patents on the process have been applied for, technical details are proprietary at this time. Clients can obtain full process information by execution of a secrecy agreement.

Preliminary estimates show that the Wellman-Lord process may have a lower capital and operating cost than the Monsanto process. These estimates also show a substantial decrease in net operating costs as coal sulfur contents increased from 3 to 4 percent.

7.1.4 Coal Refuse Compositions

Refuse composition depends upon the characteristics of the ROM coal and its response to coal preparation techniques. For the process evaluated here in Section 7 and for the one considered in Section 8, certain requirements were set regarding the clean coal and refuse from the cleaning operation. The most important criterion is that the clean coal must contain no more than 1.0 percent sulfur. Other criteria are: clean coal recovery should be no lower than 75 percent; ash content of the refuse should not exceed 50 percent.

On the basis of washability data supplied by the NAPCA for 114 coals from western Pennsylvania, eastern Ohio, and northern West Virginia, 26 coals cleanable to 1.0 percent sulfur were selected. Table A-1 summarizes the data on these coals. The groupings in this table have significance only to the Two-Stage Gasification process described in Section 8.

For use in the Combustion-Sulfur Oxides Removal process, the 26 coals in Table A-1 were used to synthesize the 10 refuse compositions shown in Table 7-1. These permitted a wide range of flexibility when used as input to the computer program for the economic study. Ranges for ash and sulfur content were set to be consistent with the washability tests and appropriate for commercial boiler operation.

Table 7-1 further shows the higher heating values on an as-fired basis (includes moisture and ash). These were calculated from the experimental value of a representative coal selected from the Keystone Coal Buyers Guide. Inspection of the data shows that about 80 percent of the total sulfur in the coal occurs as pyrite. On this basis, the mineral content was calculated by correcting the ash content for FeS_2 that was converted to Fe_2O_3 during combustion. The final heating value was calculated on the basis of corrected coal content, FeS_2 heat component, and moisture.

Table 7-1

PYRITE-COAL REFUSE COMPOSITIONS

No.	S_t % (MF)	S_p % (MF)	Ash % (MF)	Moist. %	S_t % (wet)	S_p % (wet)	Ash % (wet)	Higher Heat Value Btu/lb (wet)	Fe_2O_3 % Ash
1	7.5	6.0	30	15	6.38	5.10	25.5	8,730	25.0
2	10.0	8.0	30	15	8.50	6.80	25.5	8,665	33.3
3	5.0	4.0	40	15	4.25	3.40	34.0	7,520	12.5
4	7.5	6.0	40	15	6.38	5.10	34.0	7,455	18.7
5	10.0	8.0	40	15	8.50	6.80	34.0	7,390	25.0
6	12.5	10.0	40	15	10.62	8.50	34.0	7,325	31.2
7	7.5	6.0	50	15	6.38	5.10	42.5	6,180	15.0
8	10.0	8.0	50	15	8.50	6.80	42.5	6,115	20.0
9	12.5	10.0	50	15	10.62	8.50	42.5	6,050	25.0
10	15.0	12.0	50	15	12.75	10.20	42.5	5,985	30.0

S_t = Total sulfur
S_p = Pyritic sulfur
MF = Moisture free
Assumed ferric percentage = 20%

Basic data supplied by NAPCA

7.2 ECONOMICS

The high-specific-gravity material removed in coal preparation — coal refuse — is currently of no value. Indeed, significant costs are often incurred in its disposal. As a result, the potential value of such refuse as a raw material is difficult to estimate. Value methods which involve determination of current prices and markets for similar raw materials cannot be used without significant modifications. The properties of the coal refuse considered in this study vary too much from those of either coal or pyrite to permit meaningful comparison.

It is possible, however, to determine the probable market value of two products, electricity and sulfuric acid, produced by the Combustion-Sulfur Oxides Removal Process. Further, for set product values, it is possible to estimate the price which can be paid for refuse. This refuse price can then be used to evaluate changes in the cost of desulfurizing coal attributable to refuse processing.

7.2.1 Method of Coal Refuse Value Determination

As Figure 3-1 indicates, three separate operations — coal preparation, electric power generation, and sulfuric acid production — must be considered in evaluation of economic feasibility. If any one of these operations is uneconomical to its owner, the entire process would probably be unattractive.

The three operations could be owned and managed in a variety of ways. Where large quantities of power are to be generated, it is probable that ownership would be divided among an electric utility and one or two other organizations. When one other organization is involved, it would most probably be a chemical-energy company; when two others are involved, the coal preparation plant would probably be owned and managed by a coal or energy company and the acid recovery plant by a chemical company. The economic evaluation method used in this study is applicable to either of the foregoing types of ownership. Because of the differences between regulated (e. g. , electric utility) and nonregulated industries, this method is not applicable to instances in which all operations are owned by a single organization.

From the standpoint of the power plant owner, the energy value of the refuse is of primary importance; from the standpoint of the acid recovery plant owner, it is the sulfur value. The two values are largely independent and, where free competition exists, they can be estimated separately. If the sum of these values yields a positive market price for refuse, refuse processing is probably attractive for the coal-preparation plant owner. Conceivably, revenue from refuse sale could pay for part or all of the cost of coal preparation.

In this study, each component of refuse value (energy and sulfur) has been determined separately. This was done because of the different ways in which electric utilities and nonregulated companies view the profitability of investments and handle their accounting.

Refuse Energy Value Determination. The method used to estimate refuse energy value consisted of determining the price an investor-owned utility could pay while generating power at a competitive price. In this study, ROM coal is the fuel against which coal refuse competes. However, the general method is applicable for other fuels, both fossil and nuclear.

Two prime factors must be considered to determine the price a utility is likely to pay for coal refuse: the future load requirements of the utility and the operating costs (including fixed charges) associated with power plants firing coal refuse.

The future load requirements of the utilities are important because they relate to the capacities which are reasonable for refuse-fired power plants. To establish the optimum capacity of such plants, a compromise generally has to be reached among the load requirements of the utility; the sulfuric acid market; the low-sulfur coal market; the quantity of coal refuse available; and the economically recoverable reserves of the mine or mines supplying the coal preparation plant. The form of this compromise has a significant effect on total refuse value.

In the subject regions, the capacities of the power plants currently (1968) under construction range from 500 to 900 Mw. New plants being ordered are at the high end of this range or over it. And the trend toward larger plants will undoubtedly continue. Therefore, any refuse-fired power plant which would be built by a utility would have to be capable of competing with large-capacity units — 500 to 1000 Mw or more.

The quantity of refuse required by plants in the 500 to 1000 Mw range is large compared with the quantity currently produced by preparation plants. A 500-Mw unit would require between 260 and 420 tons per hour of refuse of the type considered in this study. This is equivalent to a preparation plant capacity of 2950 to 4580 tons per hour of ROM coal for a yield of 80 percent. For comparison, a preparation plant with a capacity of 1500 tons per hour of ROM coal is considered large. Further, the coal production required to supply refuse for a 500-Mw power plant would amount to approximately 10.0 to 15.5 million tons per year and 20.0 to 31.0 million tons per year for yields of 80 and 90 percent respectively. (In 1966, the largest mine in the nation produced only 5.65 million tons of bituminous coal.)

As the above figures indicate, matching coal mine production, coal preparation plant capacity, and the entire load requirement of a new power plant or unit would be difficult. It is possible, however, that refuse from coal preparation could be supplied from more than one mine or supplemented by reclaiming refuse from old waste. These alternatives have not been investigated; but it is doubtful that the entire fuel demand of a large capacity plant could be met by one source.

A compromise similar to that required among power generation, ROM coal production, and low-sulfur coal production would be required for sulfuric acid production. The acid production associated with a 500-Mw refuse unit would range from 980 to 3850 tons of 100-percent acid per day. The largest sulfuric acid plant in the world has a production capacity of 2000 tons of 100-percent acid per day. Thus, the maximum capacity for a refuse-fired power plant appears to be of the order of 500 Mw, a capacity smaller than that likely to be required by utilities.

In estimating refuse value, it is therefore necessary to consider the cost associated with meeting the utilities' load requirement in excess of that which can be fulfilled by a refuse plant. This is accomplished by considering two power plants or units; one refuse fired, and one non-refuse fired. The combined production and capacity-load of both plants are equal to that required by the utility.

To be economic, the dual plant scheme should produce power at the same cost as a single plant of equal capacity fired with ROM coal. Figure 7-3 shows the components of the dual plant scheme and their relationship to refuse value. The term "alternate power plant" refers to the non-refuse-fired unit in the dual plant scheme. It supplements the refuse-fired unit so that the two units match the capacity of the single plant base case.

If fuel costs were equal, the two smaller plants could not produce power as cheaply as the single large base plant. Therefore coal refuse must be bought cheaply enough to make up for the small plant size disadvantage. The step by step procedure for calculating refuse fuel value follows. Figure 7-4 outlines the steps.

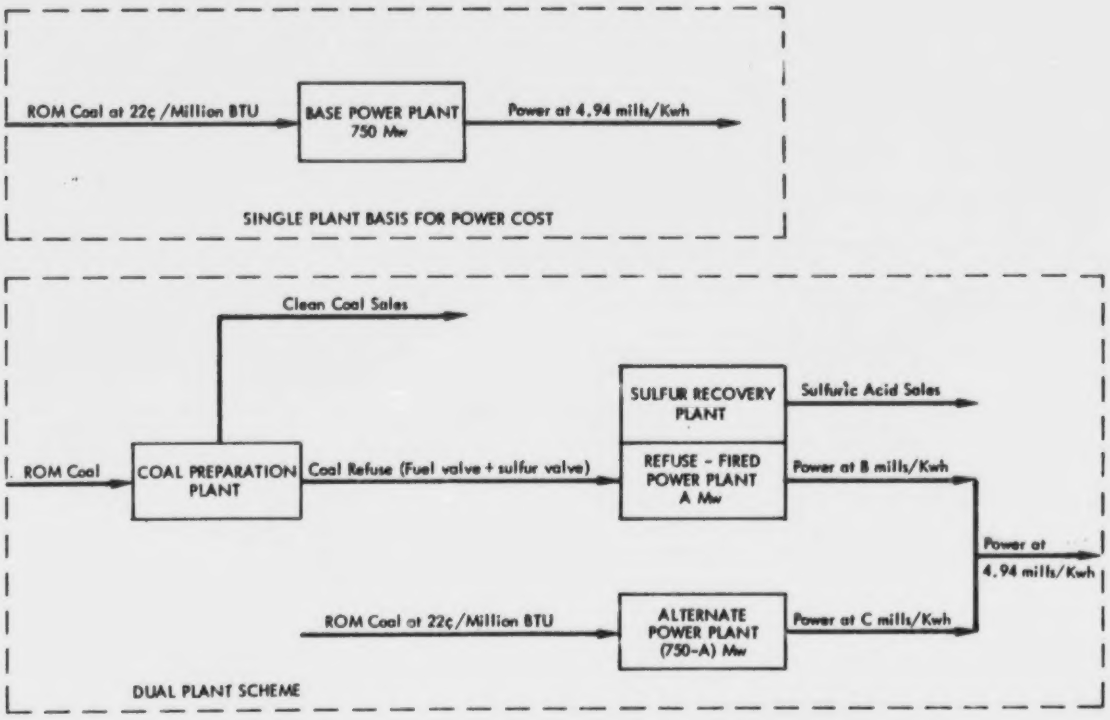

ROM Coal at 22¢ /Million BTU
BASE POWER PLANT 750 Mw
Power at 4.94 mills/Kwh
SINGLE PLANT BASIS FOR POWER COST
Clean Coal Sales
SULFUR RECOVERY PLANT
Sulfuric Acid Sales
ROM Coal
COAL PREPARATION PLANT
Coal Refuse (Fuel valve + sulfur valve)
REFUSE - FIRED POWER PLANT A Mw
Power at B mills/Kwh
Power at 4.94 mills/Kwh
ROM Coal at 22¢/Million BTU
ALTERNATE POWER PLANT (750-A) Mw
Power at C mills/Kwh
DUAL PLANT SCHEME

Figure 7-3. Dual Plant Scheme

First, the probable load required by an electric utility in the geographic region of interest is established. The costs associated with a power plant or unit capable of supplying this load are then estimated.

For most of the economics developed here, the utility load requirement was set at 750 Mw (net) and 4. 93 billion kwh per year (75 percent capacity factor). The estimated costs associated with a coal-fired power plant capable of meeting this load and their basis are shown in Appendix B.

Figure 7-4

METHOD OF ESTIMATING COAL REFUSE ENERGY VALUE

1. Establish load requirement for electric utility (base plant).
2. Estimate power cost for base plant.
3. Establish capacity of refuse-fired power plant (refuse plant).
4. Establish capacity factor of refuse plant.
5. Estimate power cost for power plant necessary to supplement refuse plant (alternate plant).
6. Estimate capital cost for refuse plant.
7. Estimate fixed charges associated with refuse plant.
8. Estimate annual operating and maintenance costs for refuse plant.
9. Calculate annual fuel cost for refuse plant.
10. Calculate refuse energy value.

A cost summary for the 750-Mw, 4. 93 billion kwh per year base plant is presented in Table 7-2. (A single, non-refuse-fired power plant, capable of meeting the specific load requirement of a utility, is referred to here as a base power plant.) This plant, like all others considered, is fueled with ROM coal. Its unit power cost is estimated to be 4. 94 mills per kwh.

After the base power plant is defined, the capacity of the refuse-fired power plant is established. Capacities of 200, 300, and 500 Mw (net) have been investigated.

Next, the refuse plant capacity factor is established. A factor of 85 percent has been used throughout. That is, all the plants are base loaded. The reasons for this are that the fuel cost (coal price) has been assumed equal for the base plant and the alternate power plant while the refuse-plant unit fuel cost (cents per million Btu) must be lower than that of the base plant. As stated before, the only way the economic advantages of the base plant can be offset by a dual plant scheme using refuse is through reducing the overall fuel cost for the scheme below that of the base plant.

Table 7-2

750-MW BASE POWER PLANT SUMMARY	
Capacity factor (percent)	75
Annual output (billion kwh)	4.93
Total capital cost (millions $)	92.2
($/kw)	122.9
Levelized Annual Cost	
Fixed charges (millions $)	13.0
Operating & Maintenance (millions $)	1.5
Fuel (millions $)	9.8
Total (millions $)	24.3
(mills/kwh)	4.94

Once the capacity and capacity factor for a given refuse plant are set, they are calculated for the alternate power plant. The costs associated with the alternate plant are then estimated. These costs and their bases for all alternate plants are presented in Appendix C.

For all alternate plants, the capital cost was determined by scaling (with a factor of 0.75) from the base plant estimated capital cost. The same data — tax rates, depreciation method, plant life, rate of return, etc. — used to estimate the fixed charges for the base plants were used for the alternate plants. And, as stated earlier, the fuel cost for the base and alternate plants were assumed to be equal. Operating and maintenance costs were estimated for each alternate plant on an individual basis.

Refuse plant capital costs were estimated next. These are shown for all plants in Part I of Appendix D. They are based on approximate prices received from steam generator vendors and on rough order of magnitude estimating for the balance of the plant. The total capital cost figure includes all direct and indirect field costs, engineering, contingency, interest during construction, and other owner expenditures. However, no allowance was made for the cost of land or land rights. Costs are based on present-day estimates and do not include escalation. The cost of the generator step-up transformers is included in the estimates, but the cost of the switchyard is not.

For each refuse-plant capacity, two estimates were prepared, one for a cyclone-fired steam generator and one for a pulverized-fuel-fired steam generator. (The technical significance of these two types of steam generators is discussed in paragraph 7.1.) The approximate prices for these two types of steam generator were received from vendors on different bases; the cyclone quotation was also unadjusted for high-ash fuels. The quotations, therefore, had to be adjusted to a common basis, and within the accuracy of the estimates it was impossible to distinguish between the cost of the two designs, except for costs attributable to ash handling — see Appendix D, Part I. For these facilities, the cyclone cost was slightly lower than for the pulverized-fuel design. As a result of the small differences between capital cost estimates for the two

Table 7-3

REFUSE POWER PLANT SUMMARY

		Capacity (Mw)		
		200	300	500
Capacity factor	(percent)	85	85	85
Annual output	(billion kwh)	1.49	2.24	3.72
Total capital cost[1]	(millions $)	37.8	51.1	74.3
	($/kw)	188.8	170.3	148.8
Levelized, annual fixed charges	(millions $)	5.3	7.2	10.5
Annual operating and maintenance costs	(millions $)	0.7	0.9	1.4
(percent of total annual cost)		11.7	10.4	8.9

1. Average value, see text

designs, and the approximate nature of the estimates, a single average capital cost was used for all economic calculations. These values are shown in Part II of Appendix D and summarized in Table 7-3.

In estimating capital costs for the refuse plants, it was impossible from available data (including that obtainable from steam generator manufacturers) to determine the quantitative effect of refuse ash or sulfur concentration. Therefore, the costs shown in Table 7-3 were used for all of the refuse compositions investigated. However, the effect upon refuse value of changes in plant capital cost were investigated for the 300-Mw plant. The findings of this investigation are presented in paragraph 7.2.2.

Refuse-plant capital cost estimates do not include any costs associated with sulfur oxides removal equipment. The method of handling these costs is discussed in the next subparagraph.

The estimates do, however, include costs associated with equipment which would normally be required to control particulate emissions. They are included in the estimated cost of the ash-handling system. Electrostatic precipitators called for under this item are, however, based on units which operate at low temperatures, around 250 F, and at efficiencies lower than those required by the Monsanto Catalytic Oxidation process. Costs associated with replacement of these units by the larger and more efficient units required by the Monsanto process have been allocated to the sulfur recovery plant. How this is accomplished and the reasoning behind it is discussed in the next subparagraph.

From refuse-plant capital cost estimates, levelized, annual fixed charges were next calculated. The data used for these calculations were consistent with those for both base and alternate power plants, and are shown in Part III of Appendix D. The actual, calculated values are shown in Table 7-3.

Annual operating and maintenance costs were then estimated for each refuse plant capacity. These are shown in Part IV of Appendix D and in Table 7-3. The operating and maintenance costs for the refuse plants are higher than those normally encountered for bituminous-coal-fired plants. The high ash content and low heating value of refuse are the principal reasons. For both cyclone and pulverized fuel plants, the quantities of coal refuse and ash which must be handled are considerably larger than necessary with coal firing. Also, refuse is more difficult material to handle. Actual quantities of refuse fired are shown in Part VI of Appendix D.

For refuse plants using pulverized-fuel steam generators, the pulverized capacity required is significantly more than for coal firing. As a result, the operating and maintenance costs associated with the pulverizing system would be higher than for coal, perhaps by as much as 100 percent for some types of refuse. With a cyclone-type steam generator, fine pulverization is not necessary, and hence the costs associated with fuel crushing would probably not be much higher than with coal. Indeed, if the "as received" refuse has a sufficiently fine consist, crushing may not be necessary with cyclone units. The influence of fuel properties upon steam generator design are discussed in detail in section 7. 1.

Furthermore, different ash-handling systems are required by the two steam generator types. In dry-bottom pulverized-fuel units, the bulk of the ash is removed from the flue gas with cyclones and electrostatic precipitators. With cyclone units, however, the ash is removed from the floor of the steam generator as molten slag. The latter type, if technically feasible, would probably feature a lower ash-handling cost, because of the high flue-gas particulate removal required by the sulfur oxides recovery system.

Even with the above differences, it was impossible to distinguish accurately between operating and maintenance costs for the two steam generator types or between refuse types. This is particularly true for fuel ash concentration as high as that of the refuse considered here. As a result, a single average annual cost was estimated for each plant capacity. These are the values shown in Table 7-3 . Also shown is the percentage of the total annual levelized cost accounted for by these operating and maintenance costs. Though they are small compared with those for annual fixed charges, they can have a significant effect upon refuse value. This effect for a 300-Mw refuse plant has been evaluated and is discussed in paragraph 7. 2. 2.

After determining all of the preceding power plant data, the levelized, annual fuel cost for the refuse plant was calculated from the following relationships:

A = B + C

where: A = base plant total annual cost
B = alternate plant total annual cost
C = refuse plant total annual cost

C = D + E + F

where: D = annual fixed charges
E = annual operating and maintenance costs
F = annual fuel cost

All the above values, except F, are calculated when plant sizes and ROM coal costs are established. Therefore, by difference:

F = A - B - D - E

All annual costs are on levelized basis.

The refuse plant levelized annual fuel cost was then converted to a year-by-year unit—cents per million Btu—fuel cost. Factors used in this conversion, refuse-plant heat rate, refuse price escalation, refuse heating value, and quantity of refuse fired, are shown in Parts V and VI of Appendix D.

Sulfur value determination: It is assumed that the owner of the sulfur recovery plant views profitability in a manner different from that of the utility owner of the refuse power plant. Therefore, a method different from that used to estimate the energy value was used to estimate the sulfur value.

The method consisted of estimating the price which the owner of the sulfur recovery plant could pay for refuse sulfur—the sulfur fraction of coal refuse—and receive a "reasonable" return on investment. The actual steps involved in applying this method are outlined in Figure 7-5.

First, for a specific power-plant capacity, the capital cost of the accompanying sulfur recovery plant was estimated. These estimates are presented in Part I of Appendix E and are summarized in Table 7-4.

All sulfur recovery plant capital cost estimates are based on exponential scaling of the estimated capital cost of a sulfur oxides removal system of the type represented by the Monsanto Catalytic Oxidation process and as applied to a conventional 1000-Mw bituminous-coal-fired power plant. Estimates based on available information indicate this cost to be between 20 and 30 million dollars. The $30 million figure has been used in this study, with a scale factor of 0.80.

Figure 7-5

METHOD OF ESTIMATING COAL REFUSE SULFUR VALUE

1. Estimate capital cost of the sulfur recovery plant for a specific refuse plant capacity.
2. Establish method of financing.
3. Establish discounted rate of return (ROI) on equity investment and calculate discounted equity investment.
4. Establish refuse type and sulfuric acid price (f.o.b., plant).
5. Calculate annual revenue.
6. Estimate annual operating cost less price paid for refuse.
7. Establish depreciation method and calculate annual income tax, other taxes, investment tax credit, interest, and debt service payments.
8. Calculate refuse sulfur value.

As for the refuse power plants, a single capital cost was determined for each sulfur plant capacity. The principal factors which prevented estimation of separate costs for each refuse composition were the limited cost and design information available in the public domain for the Monsanto process (a similar situation applies to other processes which might be applicable to refuse combustion-sulfur oxides recovery) and uncertainties in the refuse power plant design.

The design of the refuse power plant can affect sulfur plant capital cost in two basic ways. For a specific refuse composition, the design largely determines the size and efficiency of the particulate-removal equipment required, and hence its cost. The technical reasons for this are presented in section 7. 1. The other basic way in which capital costs can be affected arises from the fact that most of the major equipment in the sulfur plant is also common to the power plant. For example, the air preheater is used to heat the combustion air supplied to the steam generator and also to condense acid from combustion gases.

As a result of this duality, costs were allocated between the refuse power and acid plants on the following basis. All capital costs which would normally be required by a power plant firing refuse, but without provision for sulfur oxides removal, were charged to the power plant. Costs incurred in addition to these were charged to the sulfur plant. For instance, in the case of the particulate-removal equipment, the refuse power plant was charged with an amount estimated to be equal to the cost of the particulate-removal equipment that would be required by a similar

Table 7-4

SULFUR RECOVERY PLANT INVESTMENT

		Refuse power plant net capacity (Mw) 200	300	500
Fixed capital cost	(millions $)	8.3	11.4	17.2
Working capital	(millions $)	0.4	0.6	0.9
Interest during construction[1]	(millions $)	0.5	0.7	1.0
Total investment	(millions $)	9.2	12.7	19.1
	($/kw)	45.9	42.3	38.2
Loan[2]	(millions $)	2.6	3.6	5.4
Equity investment	(millions $)	6.5	9.0	13.6
Discounted rate of return	(percent)	12.5	12.5	12.5
Discounted equity investment	(millions $)	5.1	7.1	10.6

1. 6-percent interest for three years— period of construction
2. 30 percent of fixed and working capital

plant firing coal refuse, but not removing sulfur oxides from combustion gases. An amount equal to the difference between this cost and the estimated cost of the actual particulate-removal equipment required for the sulfur plant was allocated to the sulfur plant. This capital cost allocation procedure was applied to all equipment downstream of the point at which the combustion gases were reduced to a temperature below approximately 1000 F. This includes such equipment as economizer, electrostatic precipitators, air preheater, and induced draft fans.

After the sulfur plant capital cost was estimated, the method of financing the plant was established. For the plants investigated in this study, 30-percent debt financing was assumed, along with an interest rate of 6.0 percent. All data related to plant financing are presented in Part II of Appendix E, and total interest during construction and loan amounts are shown in Table 7-4. The effect of varying the percentage debt financing was investigated and is reported in paragraph 7.2.2.

The next step involved establishment of what constitutes for the sulfur plant owners a reasonable or attractive investment. Numerous methods which attempt to quantify the "attractiveness" of investments are presently in use. Although none of them succeeds fully in accounting for all factors, some do provide an indication of how investments compare under certain conditions. One of these methods, discounted rate of return on investment (ROI), has been used in this study to define the "attractiveness" of the sulfur recovery plant.

The ROI is the rate of interest that makes the present value of the cash proceeds from an investment (the sulfur plant) equal to the present value of the cash outlays (total equity investment) required by the investment. A specific ROI value was selected; this value was then used as the rate of interest for calculation of the present value of the equity investment required for the sulfur plant (see Part IV, Appendix E). The discounted equity investment was then used, as described below, to calculate the price which could be paid for refuse sulfur.

The refuse type and a sulfuric acid selling price to be used in a specific set of calculations were established next. FOB acid prices were assumed at \$7.50, \$10.00, \$12.50 and \$15.00 per ton of 77.7-percent acid.

From the above values, the annual revenue for the sulfur plant was calculated. Since the capacity factor for the refuse power plant was held constant, annual sulfur plant production and hence annual revenue are constant. Acid production figures for all ten refuse types and for power plants of 200, 300, and 500 Mw are shown in Appendix F.

Estimates of sulfur recovery plant annual operating costs and their bases are shown in Appendix G. The material balances used in preparing these estimates were based on a rough appraisal of the Monsanto process in its present, known state of development. Inasmuch as the design of this process may change significantly before, if ever, it achieves commercial status, the operating cost estimates presented in this study should be considered as being preliminary estimates and subject to considerable change.

For the operating cost estimates, a procedure similar to that employed for capital costs was used to allocate between the two plants the operating costs associated with common equipment. The two instances in which this procedure was used were for the electrostatic precipitator and the induced draft fans. In both cases, the total operating cost for the refuse-sulfur recovery scheme was estimated along with the cost which would be incurred by a refuse-firing power plant with no provision for sulfur oxides removal. The latter cost was allocated to the refuse power plant, and the difference between the two costs to the sulfur plant. The operating cost associated with driving the rotary air preheater was allocated entirely to the power plant, along with maintenance costs normally encountered. Other operating and maintenance costs which would be associated with a refuse plant without provision for sulfur oxides removal were also allocated to the power plant. All costs in addition to these were allocated to the sulfur recovery plant.

Sulfur recovery plant catalyst requirements were estimated from published equilibria and rate data (see Part III, Appendix G). Actual requirements, however, are subject to the type of proprietary catalyst used, and the values presented here should be regarded as being rough estimates.

In preparing the operating cost estimates, an attempt was made to include the effects of varying combustion gas/sulfur oxides concentration. Table 7-5 shows this effect for fuel types 3 through 6 for a 300-megawatt refuse power plant. Consideration has been given to changes in both combustion-gas volume and sulfur oxides concentration.

With capital costs, revenue, and operating factors determined, annual income tax, other taxes, investment tax credit, interest, and debt service payments can be calculated. The bases for these calculations are presented in Appendix H.

From the preceding data, the refuse sulfur value was calculated in the following manner:

- All components of annual cost—revenue; operating cost; income tax; investment tax credit; other taxes; interest payments; and debt service—except the unknown annual cost of refuse were discounted to their present value, using a rate of interest equal to the established ROI.
- The above present value was subtracted from the previously calculated discounted equity investment. This difference represents the present value of the cost of refuse for the sulfur plant, but uncorrected for the fact that this cost is (for income tax purposes) an expense.

Table 7-5

ANNUAL OPERATING COST FOR 300 Mw REFUSE POWER PLANT (34.0 refuse ash concentration)		
Fuel Type	Refuse Sulfur Concentration (percent, wet)	Operating Cost (thousands $)
3	4.25	1,733
4	6.38	1,873
5	8.50	1,995
6	10.62	2,106

- The above difference was corrected to account for income taxes and the actual annual value determined by "undiscounting" using a rate of interest equal to the ROI. This tax-corrected, "undiscounted" value represents the highest annual cost the sulfur plant owners can pay for the refuse and obtain the established ROI.
- From the annual value and the quantity of refuse fired by the refuse power plant, the unit refuse sulfur value—dollars/short ton—was calculated.

Total refuse value. The two refuse values—energy and sulfur—were combined to determine the total, potential value of coal refuse to the coal preparation plant owners. This value under various conditions is discussed in the following paragraph.

7.2.2 Coal Refuse Value

The method described in the preceding paragraph has been used to estimate the value of coal refuse under various conditions. From the interrelation of these values and such specific conditions as base power plant load, refuse composition, and sulfuric acid sale price, the economic limitations of refuse combustion-sulfur oxides recovery can be estimated.

Figure 7-6 shows in a graphic way the relative importance of the variables that affect total refuse value. The origin of this figure represents the specified standard conditions, the abscissa the percent change in a single, independent variable, and the ordinate the resulting incremental change in total refuse value. Although this chart was based on a 300 Mw refuse-fired power plant, the conclusions are generally true for other sizes.

The chart shows that refuse value is most sensitive to changes in acid price and capital cost of the refuse-fired power plant. For example, a 10-percent increase in acid price nearly offsets a 10-percent increase in refuse power plant investment. A more detailed analysis of refuse value and the variables that affect it follows.

Figure 7-7 relates total refuse value, refuse composition, and refuse power plant capacity for one such set of conditions: a utility load requirement of 750 Mw and 4.93 billion kwh/year, a sulfuric acid selling price of $12.50/ton, and a 12.5 percent return on sulfur recovery plant investment. The complete basis for calculation of the total refuse values shown in this figure is contained in Table 7-1 and Appendixes D through H.

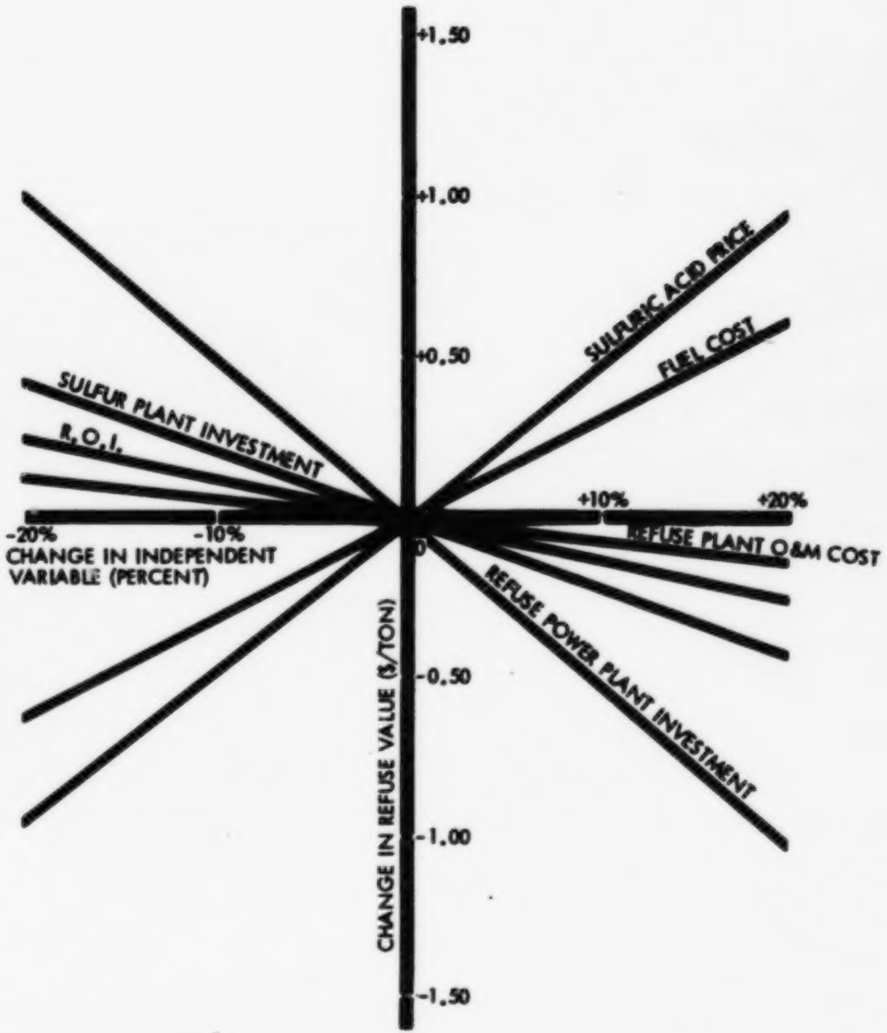


Figure 7-6. Total Refuse Value Sensitivity, 300-Mw Refuse Power Plant

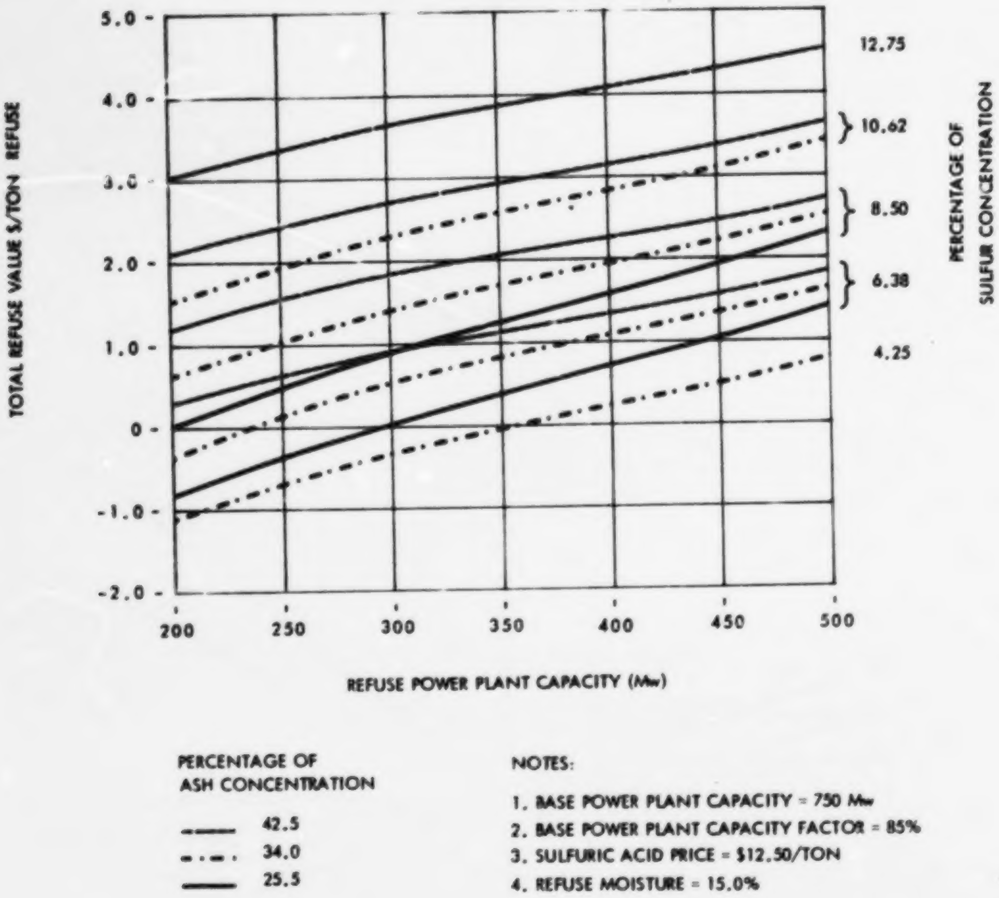


Figure 7-7. Total Refuse Value - Refuse Power Plant Capacity - Refuse Composition

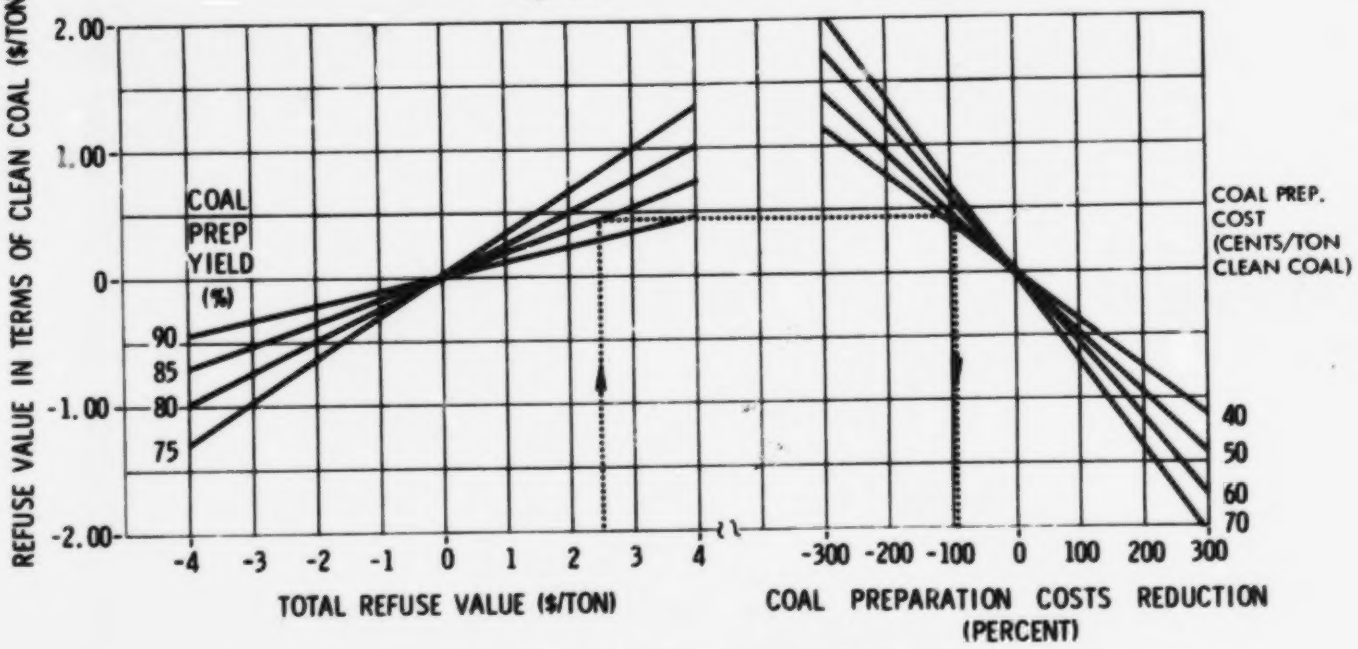


Figure 7-8. Total Refuse Valve Conversions

In Figure 7-7, the refuse values tend to be grouped according to total refuse sulfur concentration, particularly for larger power plants. Within these groups, the higher ash concentrations yield the greater refuse values. In part, this increase in value with increase in ash is explained by the greater quantity of refuse that must be fired as ash concentration increases and the fact that a single, average capital cost has been used for both refuse power and sulfur recovery plants of given capacity.

Total refuse value ($/ton) can be converted either to an equivalent clean coal value (refuse value expressed in terms of $/ton of clean coal) or to percent change in the cost of coal preparation through use of Figure 7-8. A coal preparation plant yield must be selected for conversion to an equivalent clean coal value; a yield and coal preparation cost (cents/ton of cleaned coal) must be selected for conversion to percent of change in coal preparation costs. For example, if the total refuse value equals $2.50/ton of refuse, preparation plant yield equals 85 percent, and preparation cost equals $0.50/ton of cleaned coal, then from Figure 7-8 the refuse value in terms of clean coal is $0.44/ton of clean coal and the percent of change in coal preparation cost is a reduction of 88 percent. A negative total refuse value means that the coal preparation plant owners must pay for refuse disposal; a negative value for percent of preparation cost change means that there can be a reduction in the cost of coal preparation, and a positive value means there can be an increase in cost.

Normally, the yields of coal preparation plants range between 80 and 95 plus percent. For almost all of the coals cleanable to one percent sulfur or less, the washability yields were less than 90 percent and for some coals as low as 75 percent. The estimated cost for cleaning coals similar to those investigated is on the order of $0.60/ton of clean coal.

If the $0.60/ton cost is accepted and a yield of 80 percent is assumed, the percent of change in coal preparation cost possible for the conditions which applied to Figure 7-7 are:

% of Sulfur Concentration	% of Change Over Range of Ash Concentration
4.25	between 49% increase and 32% decrease
6.38	between 36% increase and 76% decrease
8.50	between 0 and 113% decrease
10.62	between 62% decrease and 151% decrease
12.75	between 125% decrease and 188% decrease

Thus, for all sulfur concentration above 8.50 percent, regardless of ash concentration, some reduction in the cost of coal preparation can be achieved. And at the higher sulfur concentrations, all preparation costs can be offset through refuse sale.

The sulfuric acid sale price ($12.50/ton) used in calculating the values developed for Figure 7-7 and the following figures, is in terms of actual sulfuric acid concentration and not 100-percent acid. The exact concentration of the acid is unknown, but available information indicates it would probably be near 60° Bé (77.7 percent). As a result, this value has been used in determining refuse values.

The results presented in Figure 7-7 are based on evaluation of three refuse power plant capacities — 200, 300, and 500 Mw. For these plants, the effect on refuse value of variations in acid price has been investigated; and for the 300-Mw plant the effect of other independent variables has also been considered.

300-Mw Refuse Power Plant. Of all the conditions investigated in this study, those for the 300-Mw refuse-fired power plants represent the most reasonable compromise among the conflicting requirements of power production, coal preparation, and sulfuric acid production.

For a utility load requirement of 750 Mw and 4.93 billion kwh/year, the power generation balance with a 300-Mw refuse-fired and a required non-refuse-fired power plant is more reasonable than with either 200 or 500-Mw refuse-fired plants. The capacity and output (30 percent of the total requirement) of a 200-Mw refuse plant, from the viewpoint of a utility, is probably too small to be worthwhile. With a 500-Mw refuse plant, the non-refuse-fired plant is too small. (There could be other alternatives to the use of such a small alternate plant. These are discussed in the subsection dealing with 500-Mw refuse plants.)

Refuse requirements for the 300-Mw plants range from 366 to 569 tons/hour (see Appendix D). These tonnages are equivalent to coal preparation plant capacities of 1465 and 2275 tons/hour of ROM coal with an 80-percent yield. By current standards such capacities are high, but not inconceivable.

Sulfuric acid production ranges from 611 to 2392 tons/day (see Appendix F) at 77.7-percent concentration. However, only two of the ten refuse compositions, 42.5 percent ash, 10.62 percent sulfur and 42.5 percent ash, 12.75 percent sulfur, produce more than 1600 tons/day of acid. Plants of this capacity again are considered large but they are not inconceivable.

As Figure 7-7 shows, the total refuse values associated with a given refuse composition (for the 300-Mw plants) are intermediate to those for the 200 and 500-Mw plants. The results of an evaluation of the sensitivity of refuse value to changes in specific independent variables are approximately valid for the other plants.

The effect upon total refuse value of variations in sulfuric acid sale price is shown in Figures 7-9, 7-10, and 7-11 Because of the importance and instability of this variable, its effect is shown for all three refuse ash concentrations investigated, and for a range of sulfur concentrations. The conditions which apply to the refuse values shown in these figures are specified in Table 7-6.

For a constant acid price, refuse value increases with either ash or sulfur concentration. As sulfur concentration increases, the spread increases between different acid prices and refuse values. Thus, a change in acid price results in a greater net change in refuse value for "high sulfur" refuse than for "low sulfur" refuse.

The relation between refuse value and refuse sulfur and ash concentration is shown more clearly in Figure 7-12. At a constant acid price of $12.50/ton, the sulfur concentrations at which refuse value equals zero are: 3.85, 5.10, and 6.35 percent, for ash concentrations of 42.5, 34.0, and 25.5 percent respectively.

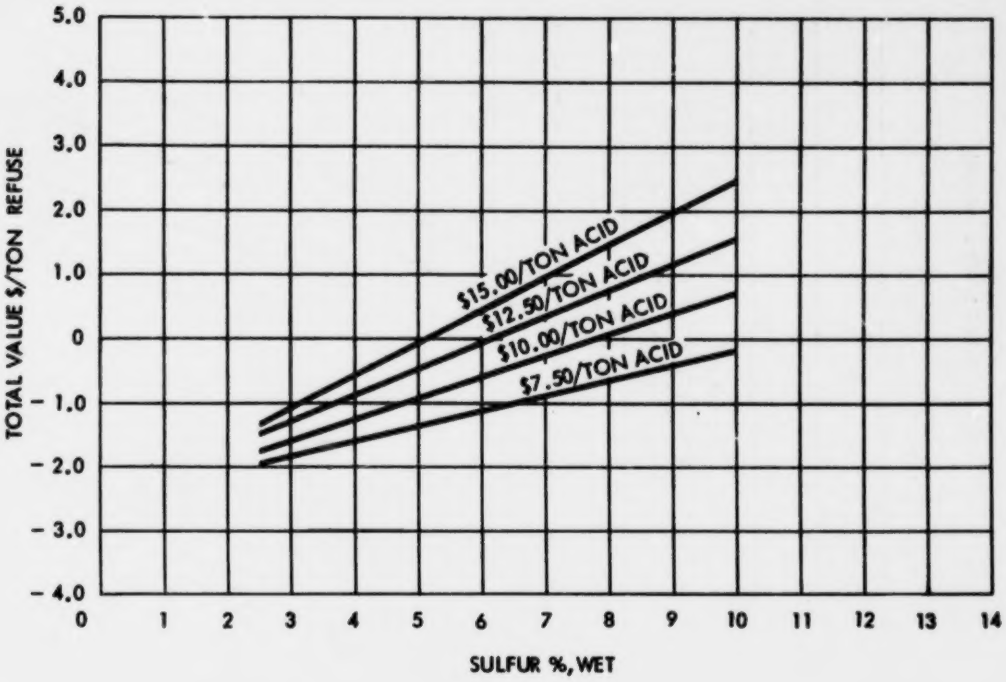


Figure 7-9. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price (Refuse Ash Concentration, 25.5%)

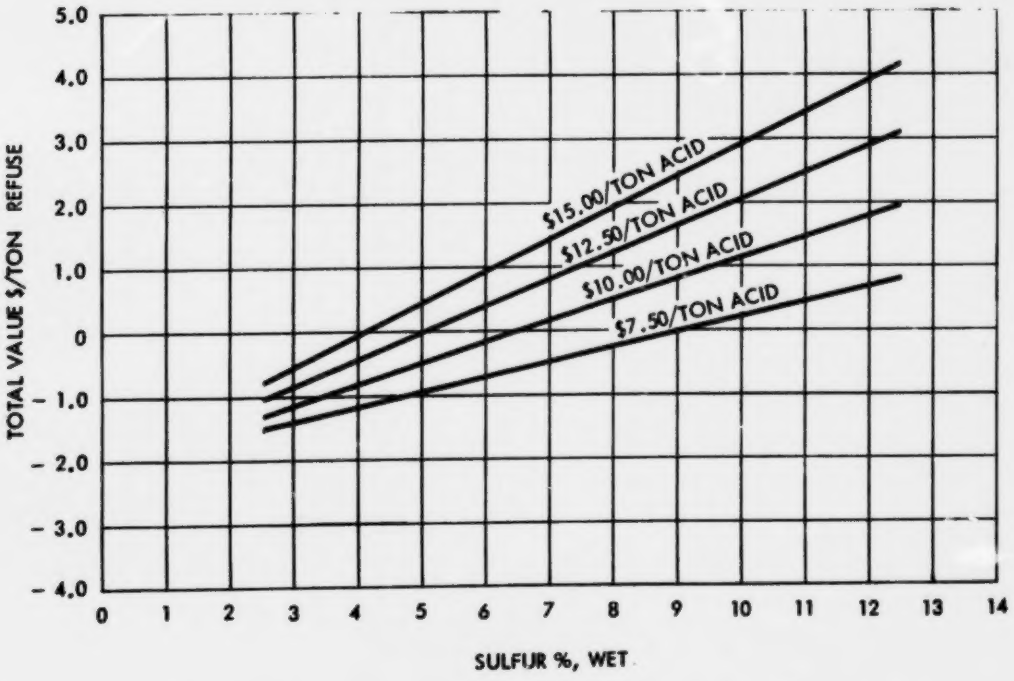


Figure 7-10. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price (Refuse Ash Concentration, 34.0%)

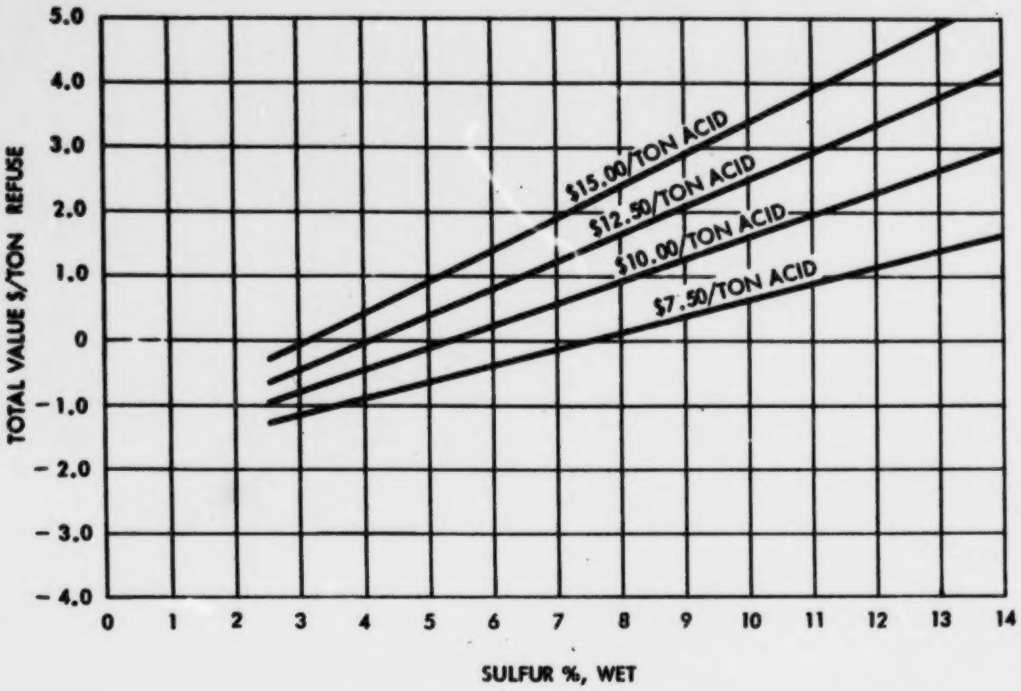


Figure 7-11. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price (Refuse Ash Concentration, 42.5%)

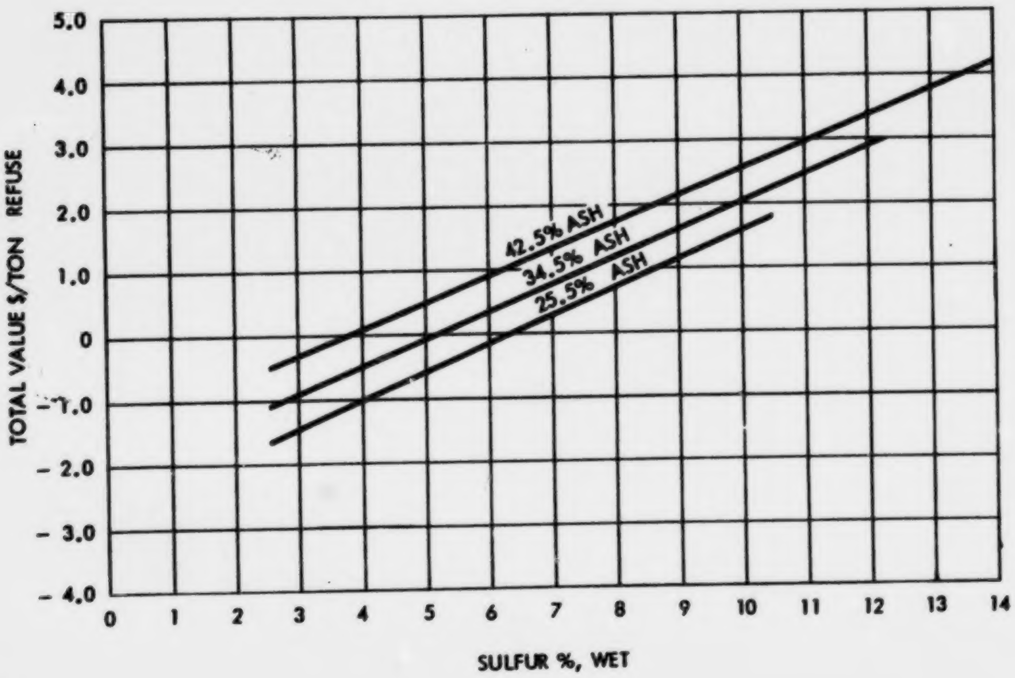


Figure 7-12. Total Refuse Value - Refuse Sulfur Concentration - Refuse Ash Concentration (Sulfuric Acid Price, $12.50/ton)

Table 7-6

STANDARD CONDITIONS
300-MW REFUSE POWER PLANTS

I.	**Base Power Plant**		
	Net capacity	750	Mw
	Capacity factor	75	percent
	Annual output	4.93	billion/kwh
	First year fuel cost	19.8	cents/million Btu
	Fuel cost escalation	1.5	percent/yr.
	Full specification: see Appendix B, 750 Mw Plant		
II.	**Alternate Power Plant**		
	Net capacity	450	Mw
	Capacity factor	68	percent
	Annual output	2.70	billion/kwh
	First year fuel cost	19.8	cents/million Btu
	Fuel cost escalation	1.5	percent/yr.
	Full specification: see Appendix C, 450 Mw Plant		
III.	**Coal Refuse Power Plant**		
	Net capacity	300	Mw
	Capacity factor	85	percent
	Annual output	2.24	billion kwh
	Total capital cost	$51.1	million
		170.3	$/kw
	Full specifications all refuse compositions: see Appendix D, 300 Mw Plants		
IV.	**Sulfur Recovery Plant**		
	Total investment	$12.7	million
		42.3	$/kw
	Discounted rate of return	12.5	percent
	Debt financing	30	percent
	Full specifications all refuse compositions: see Appendix E, 300 Mw Refuse Power Plants		

A zero refuse value implies that the coal preparation plant owners receive zero revenue for their refuse. Even so, an advantage accrues because the cost normally associated with refuse disposal is no longer incurred. Furthermore, regardless of how this condition is viewed by owners of the coal preparation plant, at zero value the owners of both the utility and the sulfur recovery plant receive a return on their investments. Under the conditions shown in Table 7-6, this amounts to 8.0 percent on book value for the utility and 12.5 percent on discounted equity investment for the sulfur recovery plant.

For evaluation of the effects of other independent variables, a slightly different technique has been used. It differs from that used for acid price only in that a single refuse ash concentration, 34.0 percent, has been considered. As Figure 7-7 shows, for a given sulfur concentration the total refuse values for this ash concentration are between those for 25.5 and 42.5 percent.

The sensitivity of refuse value to these variables has been evaluated by changing their values one at a time and comparing the resulting change in total refuse value with a standard set of conditions. The standard conditions which have been used are those specified in Table 7-6 and an acid price of $12.50/ton. These conditions are represented by the 34.0 percent ash curve in Figure 7-12 and the $12.50/ton acid price curve in Figure 7-10. The components of the refuse values for these conditions are indicated in Figure 7-13.

Refuse value sensitivity to one of the most difficult variables, refuse power plant capital cost, is shown in Figure 7-14. Four capital costs are shown in addition to the standard cost of $170/kw. These represent increases and decreases of 10 and 20 percent. A 10-percent change in capital cost yields a $0.50/ton change in refuse value. The incremental change in refuse value for a given change in capital cost is essentially

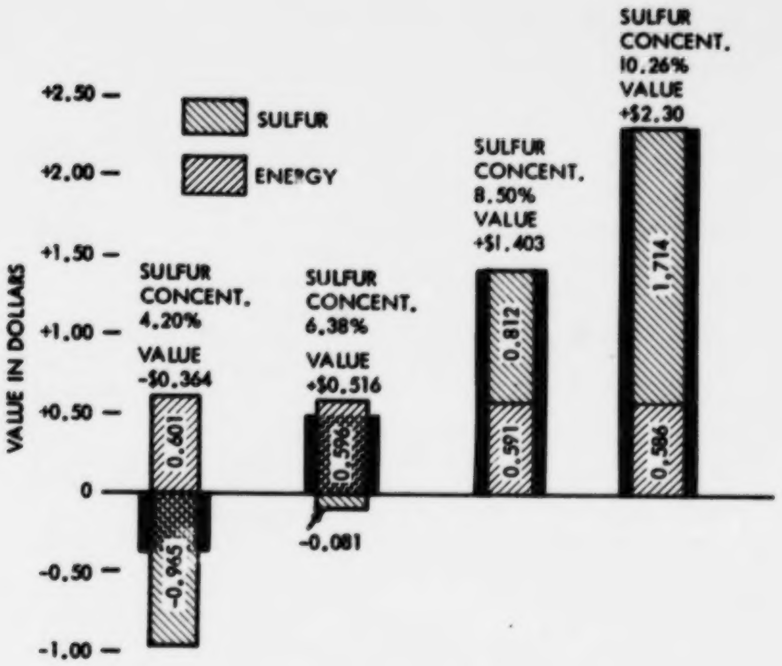


Figure 7-13. Total Refuse Value Components (300-Mw Refuse Power Plant; Refuse Ash Concentration, 34.0%)

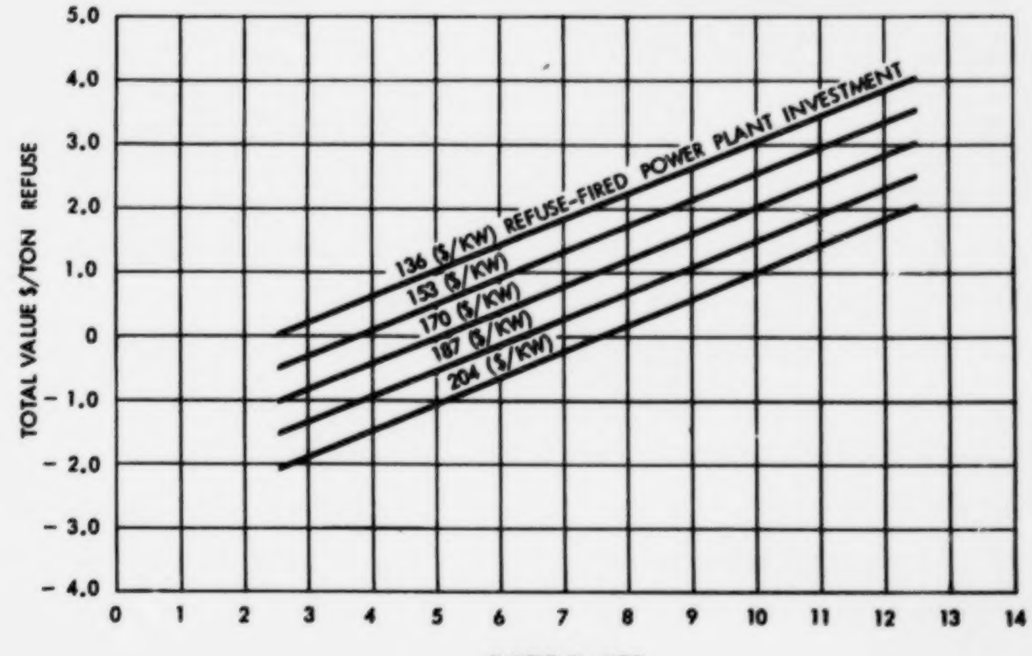


Figure 7-14. Total Refuse Value - Refuse Sulfur Concentration - Refuse Power Plant Capital Costs (Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

independent of changes in refuse sulfur concentration, at least over the range of sulfur concentrations investigated. The effect of a change in power plant capital cost upon refuse value is probably shown more clearly in Figure 7-6.

In calculating the refuse values shown in Figures 7-9 through 7-12, the same refuse power plant capital cost was used for all refuse compositions (see Refuse Value Energy Determination in paragraph 7.2.1). The assumption that capital cost is independent of refuse composition is unrealistic, but in the light of available information it is the most reasonable that could be made. The relationship shown in Figure 7-6 can be used to judge the possible consequences of this assumption. A 10-percent increase in capital cost is roughly equivalent to an incremental increase in refuse sulfur concentration of 1.3 percent.

The effect of changes in operating and maintenance cost of the refuse power plant has been evaluated in the same manner as for capital costs. As shown in Figure 7-6, refuse value is considerably less sensitive to this variable. A 10-percent change yields a $0.07/ton change in refuse value.

Sulfur recovery plant investment is another quite important and difficult variable. The relation between it and refuse value is shown in Figure 7-15. Here, investment has been expressed in terms of refuse power plant capacity, the standard value being $42/kw. The other values represent a decrease of 33 percent, $23/kw, and decreases and increases of 10 and 20 percent. Refuse value sensitivity is shown in Figure 7-6. A 10-percent change in sulfur plant investment produces a $0.21/ton change in refuse value.

The sensitivity of refuse value to changes in both refuse power plant capital cost and sulfur recovery plant investment for a sulfur concentra-

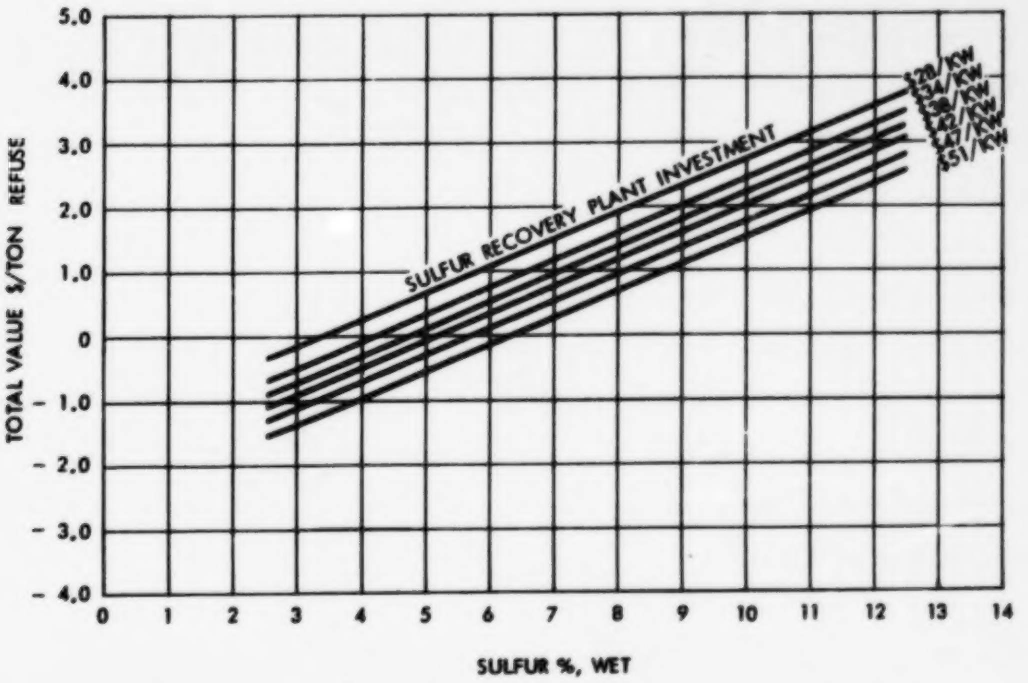


Figure 7-15. Total Refuse Value - Refuse Sulfur Concentration - Sulfur Recovery Plant Investment (Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

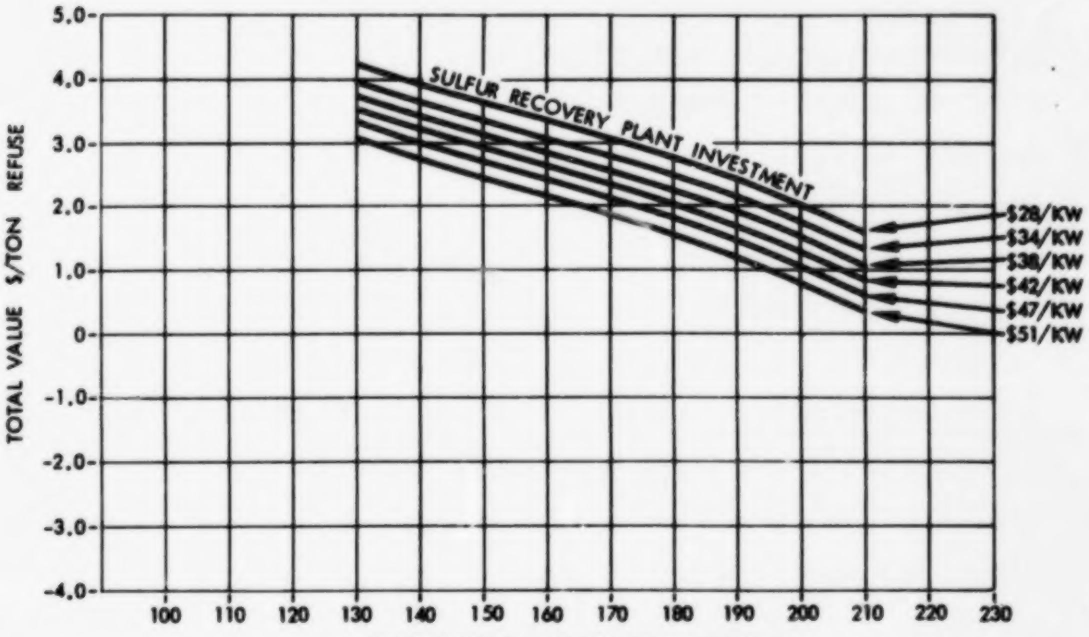


Figure 7-16. Total Refuse Value - Refuse Power Plant Capital Cost - Sulfur Recovery Plant Investment (Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

tion of 10.62 percent is shown in Figure 7-16. If each of these variables could be reduced by 20 percent from the selected standard conditions of $170/kw and $42/kw, refuse value would increase by $1.43/ton. If both variables could be increased by 20 percent, refuse value decreases by $1.50/ton.

Two variables, other than investment, important to the owners of the sulfur recovery plant are the discounted rate of return on equity investment and percent debt financing. Figure 7-17 relates the effect of a 20-percent change in ROI. An increase in ROI to 15.0 percent decreases refuse value by $0.26/ton and a decrease in 10.0 percent increases it by $0.26/ton.

The difference between zero and 30 percent debt financing is shown in Figure 7-18. Use of 30-percent financing with 6.0 percent interest results in an increase in refuse value of $0.28/ton.

For the base power plant, the independent variable subject to the greatest change is probably fuel cost. The effect of changes in this variable are shown in Figures 7-19 and 7-6. A 20-percent increase over the standard first year of fuel cost of 19.80 cents/million results in a $0.62/ton increase in refuse value. A 20-percent decrease results in a $0.62/ton decrease in refuse value.

200-Mw Refuse Power Plant. For 200-Mw refuse power plants, only the effect of the acid price upon refuse value has been investigated. Results are shown in Figure 7-20 through 7-22. The basis for these results is provided in Appendixes B through H.

Comparison of the results with those for the 300-Mw plants, Figures 7-9 through 7-11, reveals that, for a given refuse composition and acid price, refuse value is lower for the 200-Mw plants. The actual

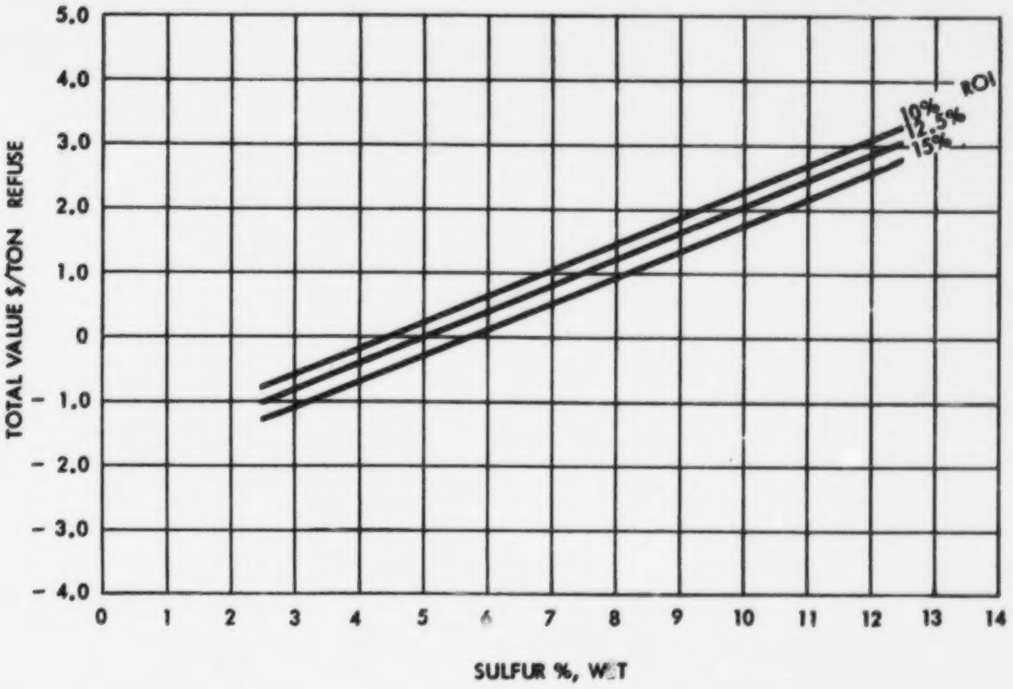


Figure 7-17. Total Refuse Value - Refuse Sulfur Concentration - Rate of Return on Investment (Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

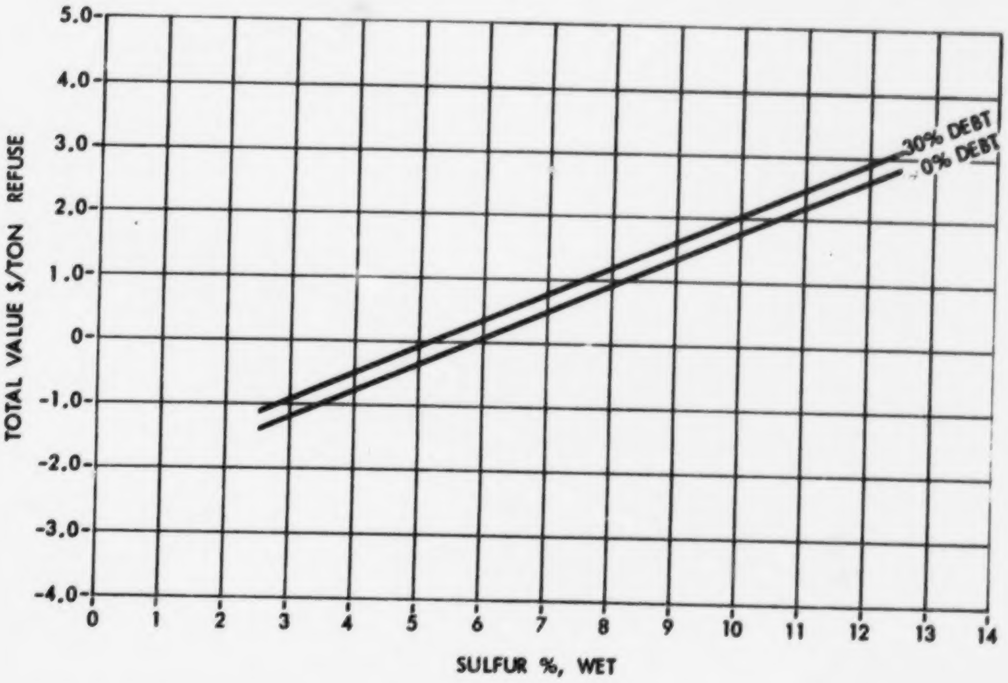


Figure 7-18. Total Refuse Value - Refuse Sulfur Concentration - Percent Debt Financing Sulfur Plant (Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

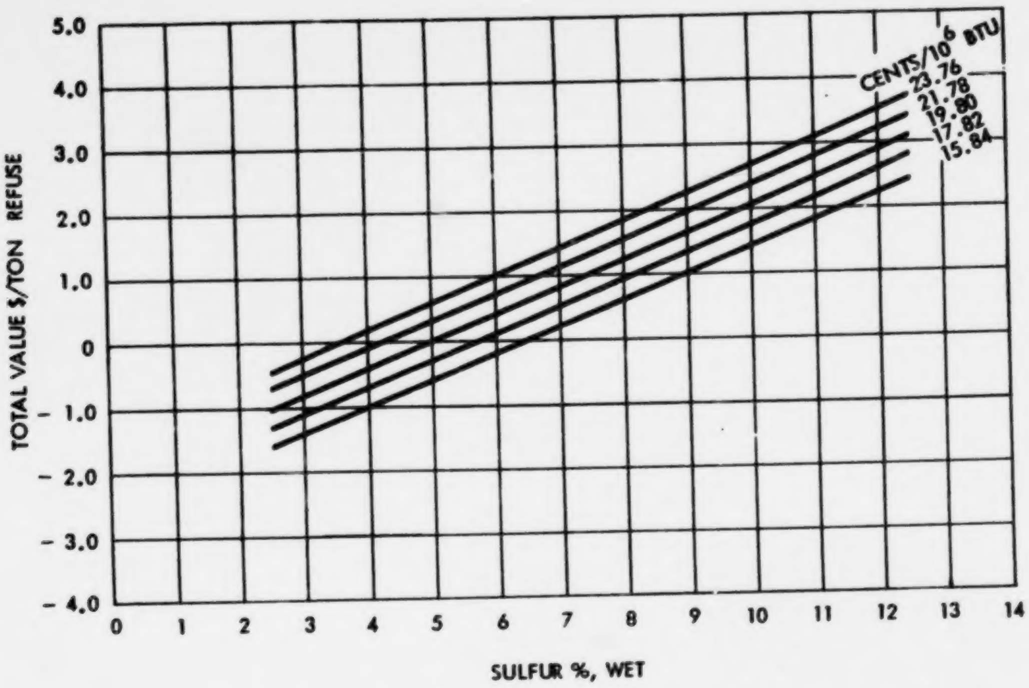


Figure 7-19. Total Refuse Value - Refuse Sulfur Concentration - Base and Alternate Power Plant First-Year Fuel Cost (1.5%/year Fuel Cost Escalation; Refuse Ash Concentration, 34.0%; Sulfuric Acid Price, $12.50/ton)

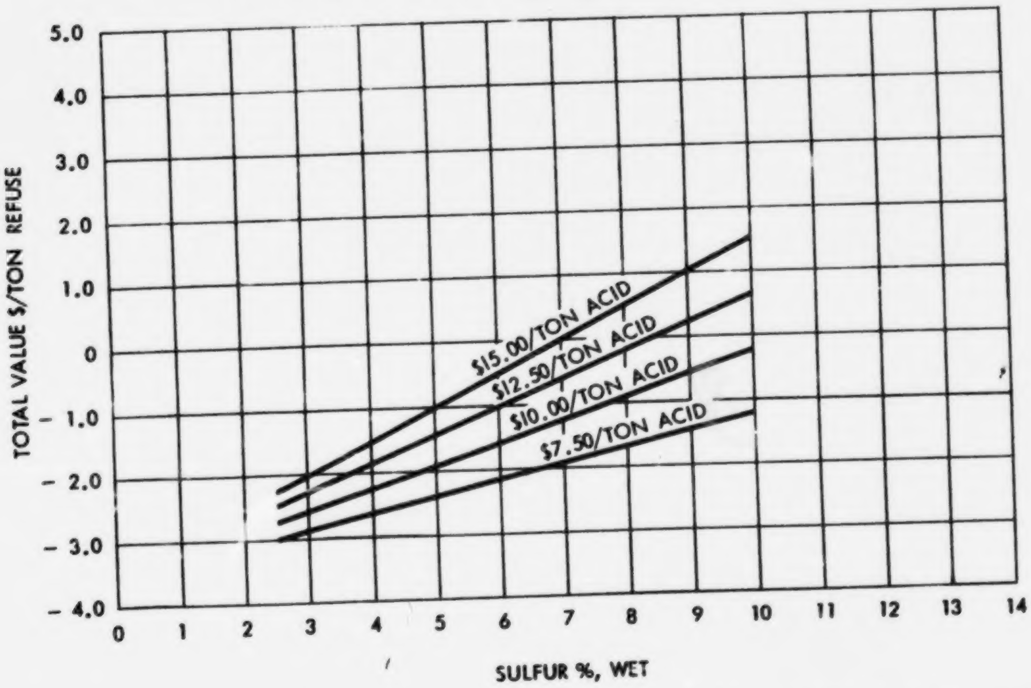


Figure 7-20. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price: 200-Mw Plant (Refuse Ash Concentration, 25.5%)

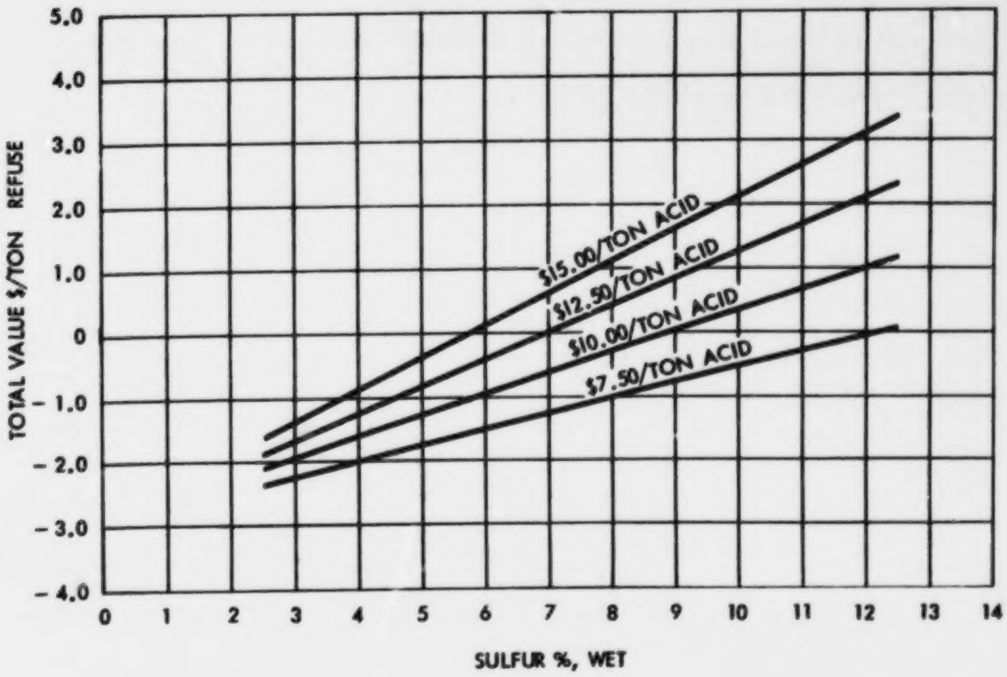


Figure 7-21. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price: 200-Mw Plant (Refuse Ash Concentration. 34.0%)

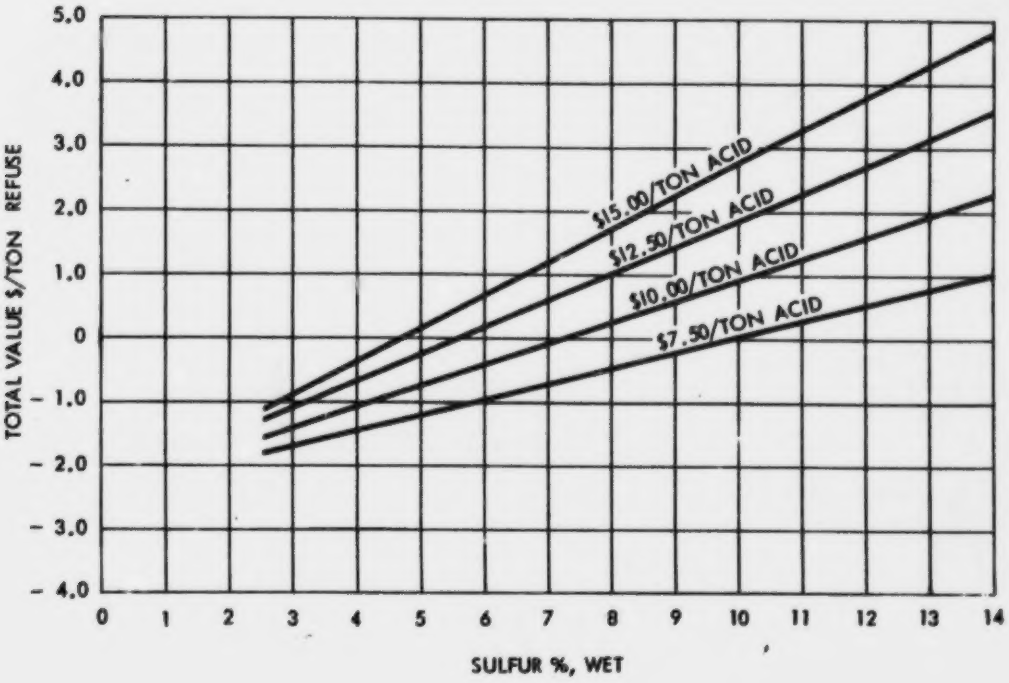


Figure 7-22. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price: 200-Mw Plant (Refuse Ash Concentration, 42.5%)

increase in value is a function of refuse sulfur concentration. At 34.0 percent ash and 8.5 percent sulfur, the difference between the two plant capacities amounts to $0.81/ton.

500-Mw Refuse Power Plant. Under the utility load requirement used for the 200 and 300-Mw refuse plants, the capacity of the alternate plant required to supplement a 500-Mw refuse power plant amounts to 250 Mw. A conventional plant of this capacity is quite small by present-day utility standards especially if it is part of a scheme involving a 500-Mw power plant. Therefore, to make the case of the 500-Mw refuse plant more realistic, at least from the utility's viewpoint, the utility load requirement was increased to 1000 Mw and 6.57 billion kwh/year. This increases the capacity of the alternate plant to 500 Mw.

Refuse values for these conditions and a refuse ash concentration of 34.0 percent are shown in Figure 7-23. The complete basis for calculation of these values is contained in Appendixes B through H.

The capacity limitation of the 500-Mw case might also be overcome by purchase of the required supplemental power from other utilities or by uprating the capacity of future plants. Neither of these alternatives has been investigated, but it is doubtful that they would change dramatically the results in Figure 7-23.

The values shown in Figure 7-23 are not the same as those of Figure 7-7. In order to be consistent, the latter values are based on a 750-Mw base power plant.

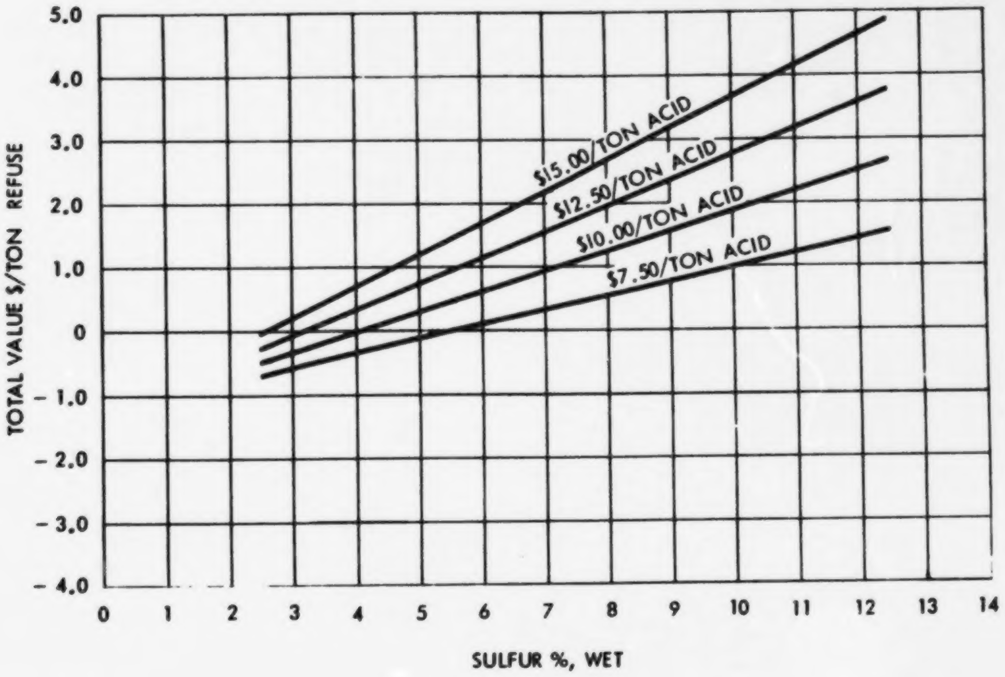


Figure 7-23. Total Refuse Value - Refuse Sulfur Concentration - Sulfuric Acid Price: 1000-Mw Plant and 500-Mw Refuse Power Plant (Refuse Ash Concentration, 34.0%)

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Section 8

TWO-STAGE GASIFICATION

To the best of Bechtel's knowledge, the gasification procedure described here is an original approach. It differs significantly from single-stage partial combustion of coal and is specially conceived for treating pyrite-coal. The study shows that gasification of washed refuse containing less than 50 percent ash is technically feasible. This process is in the conceptual stage and requires design of a suitable furnace. However, the ideas described here could become the basis for developing a process design, economic study, and pilot plant program.

Gasification, or incomplete combustion, can produce a gas in which the sulfur occurs mostly as hydrogen sulfide that can then be treated to recover elemental sulfur. The standard gas producer uses countercurrent flow of coal and moist air, giving a gas with a fuel value of about 150 Btu per cubic foot from which tar condenses on cooling. This partial combustion also leaves powdery ashes containing about 15 percent unburned carbon.

Such a system would not be suitable for refuse treatment even if its use were not precluded by the fine sizing of the refuse. Loss of carbon amounting to 15 percent of ash weight would markedly decrease the fuel available; a powdery ash would present a major disposal problem; and by-product tar would not only require special treatment but would also complicate gas cleaning ahead of sulfur recovery. With fine feed, the gasifier must operate with concurrent flow and, to minimize carbon loss and simplify disposal of residue, the ash should be fused.

Iron from the pyrite would undoubtedly lower the fusion point of the ash, but no precise data are available.

8.1 TWO-STAGE FURNACE CONCEPT

Simple partial combustion would produce gas that contained only the organic sulfur from coal and the labile sulfur from pyrite; ferrous sulfide would not be oxidized, but would report with the slagged ash. This problem could probably be solved by employing a vertical cyclone furnace equipped with two burner ports, one above the other. Refuse would be burned at the upper port with a deficiency of air to produce a reducing gas and slag, containing ferrous sulfide, which would run down the furnace walls into the flame from the lower port. At the lower level, refuse would be burned with an excess of air to ensure both complete combustion of the refuse and oxidation of the ferrous sulfide from above. Clean slag, containing about two percent sulfur, would flow to a tap in the bottom of the furnace; the gas would pass into the upper section where it would react with the reducing gas to produce an H_2S-rich gas of the composition desired for subsequent treatment.

8.2 PROCESS DESCRIPTION

Figure 8-1 illustrates a gasifier, fired with coal refuse, operating in a system that converts the fuel value to electricity and recovers elemental sulfur from the gas.

Hot gases, which contain most of the sulfur as H_2S, leave the top of the gasification furnace, generate steam in a heat recovery boiler, then preheat air for combustion of these gases later in the system.

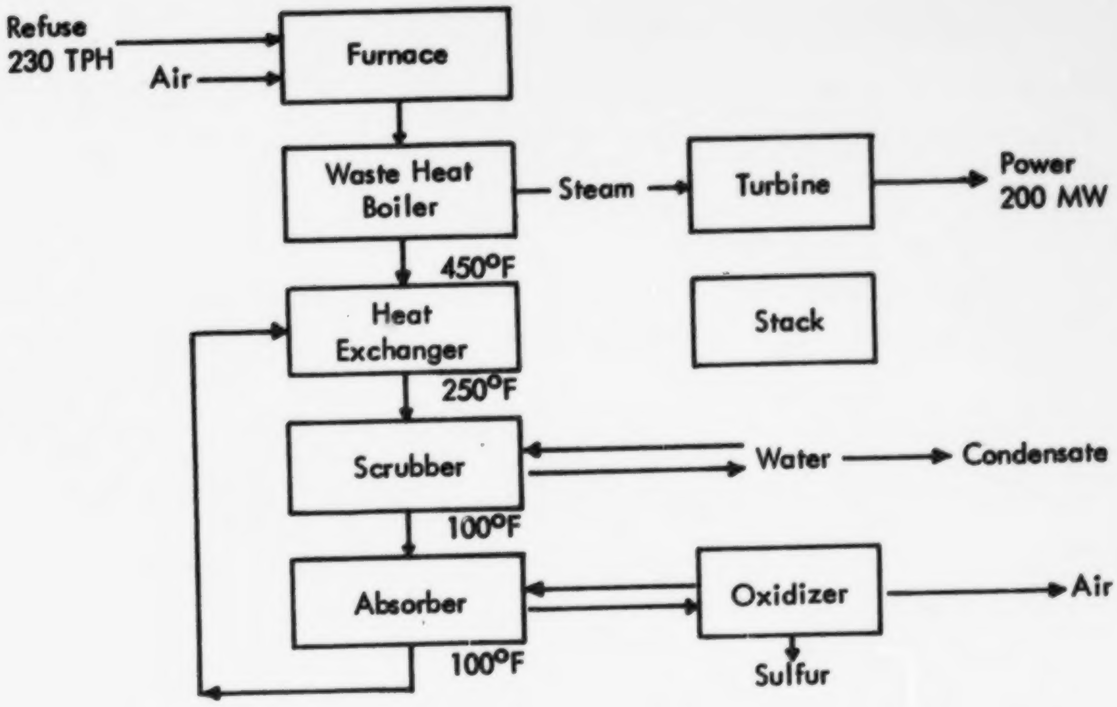
Refuse
230 TPH
Air
Furnace
Waste Heat Boiler
Steam
Turbine
Power
200 MW
450°F
Heat Exchanger
Stack
250°F
Scrubber
Water
Condensate
100°F
Absorber
Oxidizer
Air
100°F
Sulfur

Figure 8-1. Combustion of Group B Refuse

Gas, now at 250 F, is scrubbed with 100 F water to remove dust and condense water vapor. Hydrogen sulfide is removed from the gas by an absorbent solution and then converted to elemental sulfur by a Claus kiln or perhaps the Giammarco-Vetrocoke process. For lack of data, no provision was made in this study for the presence of organic sulfur components in the gas stream. Carbonyl sulfide (COS), for example, might be formed in small amounts, and then the addition of steam to the gas in contact with a catalyst would be necessary to form hydrogen sulfide by hydrolysis. Clean gas from this system would contain about 1.0 grain of hydrogen sulfide per cubic foot, and sulfur recovery would be about 98 percent of that absorbed. The sulfur would probably be contaminated with small amounts of arsenic from the absorbent solution, but purification with lime water, as practiced at Rio Tinto, Spain, would be practical.

The clean gas from the sulfur recovery steps would be burned with the preheated air from the heat exchanger in a boiler equipped for preheating combustion air for the gasifier. Waste gas at about 450 F would pass to a stack. Steam from this boiler and from the heat recovery boiler would be available for electrical power generation.

8.3 CALCULATIONS

Table 8-1 lists process stream quantities and gas compositions for the two-stage gasifier feeding coal refuse. The scale of operation is based on producing enough fuel gas to generate 200 megawatts of electric power. Calculations were made using three different refuse compositions designated as A, B, and C which are further defined in Table 8-2 and Table A-1 in Appendix A.

Coal in the refuse was assumed to contain 85 percent carbon, 5.7 percent hydrogen, 7.5 percent oxygen, and 1.6 percent nitrogen. Preliminary estimates indicate that clean gas from refuse of Groups B and C would have a maximum calorific value of 80 Btu per cubic foot; all calculations were made on this basis, although the higher coal content of Group A refuse would permit production of 100 Btu gas.

Stream quantities that appear on Figure 8-1 are based on refuse from Group B.

Other assumptions used in the calculations are:

- Because of the high ash content of the refuse, combustion air is preheated to 800 F.
- The temperature of the slagged ash is 2200 F and its heat content is 1000 Btu per lb.
- Heat supplied to both boilers is converted to steam at 85 percent efficiency which is then used at 32 percent efficiency to generate power.

Table 8-1

GASIFICATION OF REFUSE
(Power Output, 200 Megawatts)

GROUP	A	B	C
Dry Refuse, T/hr	160	220	250
Air scfh (70 F)	20.3×10^6	20.1×10^6	19.9×10^6
Slag, T/hr	35	85	100
Raw Gas, scfh (70 F)	25.5×10^6	25.4×10^6	25.3×10^6
CO %	18.9	18.9	18.9
CO_2 %	6.3	6.3	6.3
H_2 %	5.9	5.7	5.8
H_2O %	4.8	4.6	4.7
H_2S %	0.9	1.7	1.8
N_2 %	63.2	62.8	62.5
Sulfur Recovered, T/hr	9.2	17.7	18.4
Clean Gas, scfh (70 F)	24.1×10^6	23.9×10^6	23.7×10^6
CO %	20.0	20.2	20.1
CO_2 %	6.6	6.7	6.8
H_2 %	6.2	6.1	6.2
H_2O %	0.2	0.2	0.2
H_2S %	0.01	0.01	0.01
N_2 %	67.0	66.8	66.7
Air, scfh (70 F)	18.8×10^6	18.7×10^6	18.6×10^6
Waste Gas, scfh (70 F)	39.7×10^6	39.2×10^6	38.9×10^6
CO_2 %	16.2	16.3	16.3
H_2O %	3.9	3.6	3.6
O_2 %	2.0	2.0	2.0
SO_2 %	0.01	0.01	0.01
N_2 %	77.9	78.1	78.1

Table 8-2

REFUSE COMPOSITIONS (FEEDS TO GASIFIER)

COAL GROUP	A	B	C
Float Test Sp. Gr.	1.30	1.40	1.90
ROM Coal			
% S_t	3.83	2.49	2.21
% S_p	2.65	1.72	1.51
% Ash	12.45	11.04	12.47
Float (clean coal)			
% Wt.	48.1	80.1	91.1
% S_t	0.97	0.92	0.81
% S_p	0.26	0.24	0.30
% Ash	2.76	4.28	6.71
Sink (refuse)			
% Wt.	51.9	19.9	8.9
% S_t	6.35	9.08	17.04
% S_p	4.86	7.64	14.20
% Ash	21.62	38.74	71.41
Refuse Composition			
% Mineral Matter	26.8	46.8	86.5
% Coal	73.2	53.2	13.5

Notes:

S_t = total sulfur

S_p = pyritic sulfur

% Mineral matter = 1.08 x % ash + 0.55 x % S_t

Table 8-3

GASIFICATION OF REFUSE (ALTERNATIVE PROCESS)
(Power Output, 200 Megawatts)

Dry Refuse, T/hr	230
Air, scfh, (70 F)	34.0×10^6
Slag, T/hr	90
Raw Gas, scfh (70 F)	36.8×10^6
CO_2 %	17.3
CO %	0.9
H_2 %	1.5
H_2O %	5.9
H_2S %	1.3
COS %	0.2
N_2 %	73.0
Sulfur Recovered, T/hr	18.4
Clean Gas, scfh (70 F)	34.2×10^6
CO_2 %	18.6
CO %	1.0
H_2 %	1.6
H_2O %	0.2
H_2S %	0.01
N_2 %	78.6

- The refuse is dried from 15 to 5 percent moisture before gasifying. About 350, 000 Btu per ton dry refuse was deducted from the heat available for steam production, although this work would probably be done by waste steam from the turbines.

8.4 ALTERNATIVE PROCESS

Although relatively simple, the gasification process requires two boiler installations as well as a system for sulfur recovery. Thermodynamic data indicate that refuse combustion could proceed nearly to completion without destruction of the hydrogen sulfide in the gas stream. If this were possible, the cost of the second boiler could be saved at the expense of a small loss of unburned gas, and sulfur could still be removed from the gas stream. It is recognized that equilibrium would not necessarily be achieved in the combustion products, but adjustment of gas composition by catalysts appears possible.

Figure 8-2 illustrates this alternate process, and Table 8-3 summarizes the results of the calculations.

For the purposes of computation, the air supplied with the refuse was assumed arbitrarily to be sufficient to burn all the carbon and half the free hydrogen in the coal. Free hydrogen was defined as that remaining after deducting an amount sufficient to combine with oxygen in the coal and sulfur, both organic and pyritic, in the refuse.

In view of the higher furnace temperatures to be expected, the air was not assumed to be preheated. Gas from the boiler at 450 F would be cooled to 250 F in a heat exchanger which would heat clean gas and permit its disposal via a stack.

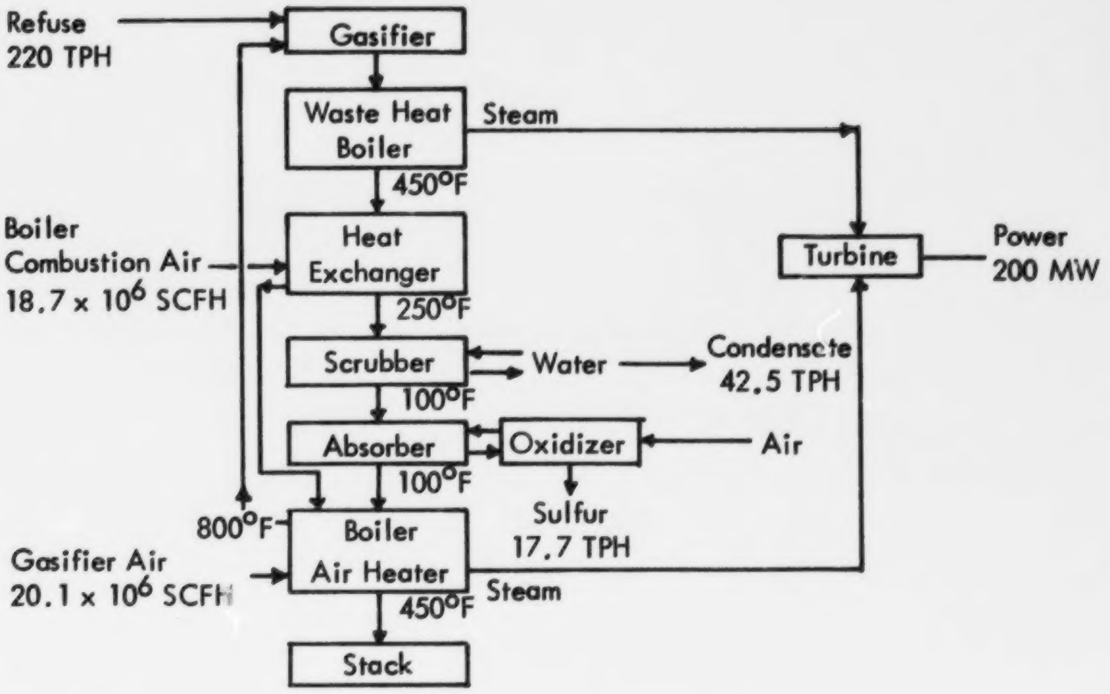
Refuse
220 TPH
Gasifier
Waste Heat Boiler
Steam
450°F
Boiler Combustion Air
18.7 x 10^6 SCFH
Heat Exchanger
250°F
Turbine
Power
200 MW
Scrubber
Water
Condensate
42.5 TPH
100°F
Absorber
Oxidizer
Air
100°F
Sulfur
17.7 TPH
800°F
Boiler Air Heater
Gasifier Air
20.1 x 10^6 SCFH
Steam
450°F
Stack

Figure 8-2. Gasification of Group B Refuse

A calculation based on refuse from Group B indicated that about 90 percent of the fuel value of the refuse would be released by combustion under these conditions, and that the waste gas would contain 2.6 percent unburned carbon monoxide and hydrogen. In view of the speculative nature of the calculation, the possibility of catalytic oxidation of the unburned gas before discharge to the atmosphere was not investigated.

APPENDIX A

REFUSE COMPOSITION

For the gasification process, the 26 coals of Table A-1 were used as classified therein into 5 groups on the basis of coal recovery and refuse composition. For simplicity, only data for coal in the 14 mesh x 0 size range were considered.

- Group A-These coals could be cleaned to one percent sulfur only at low specific gravity, and so both coal recovery and the ash content of refuse were low. Refuse treatment, therefore, would be essential to economic sulfur control.
- Group B-In these tests, coal recovery was over seventy percent, and the refuse contained less than fifty percent ash.
- Group C-At specific gravity of 1.90, the refuse from these coals contained more than fifty percent ash, and coal recovery was over eighty percent. Sulfur in the refuse, however, was high, and cleaning at a lower specific gravity might be warranted if this sulfur could be recovered.
- Group D-The results from coals in this group were similar to those in Group C except that sulfur in the refuse was low. Washing at a lower gravity, and treatment of the refuse, would probably be warranted only if the value of the cleaned coal were enhanced by lower sulfur and ash.
- Group E-These coals, all of which contained high ash, could not be cleaned with high recovery of coal, yet the ash content of the refuse was high. Treatment of such refuse would not be feasible.

Table A-1

CLASSIFICATION OF COALS

STATE	FLOAT TEST Sp. G.	COAL			CUM. FLOAT				CUM. SINK			
		% S_t	% S_p	%Ash	% Wt.	% S_t	% S_p	%Ash	% Wt.	%S_t	%S_p	%Ash
GROUP A												
OHIO	1.30	4.85	3.52	11.47	47.7	0.89	0.24	2.18	52.3	8.47	6.53	19.94
PENNA	1.30	3.36	2.29	14.32	46.9	1.04	0.37	3.98	53.1	5.41	3.99	23.45
OHIO	1.30	2.39	1.39	9.97	58.3	1.03	0.16	2.35	41.7	4.29	3.24	20.62
OHIO	1.30	4.73	3.53	14.06	39.6	0.91	0.26	2.54	60.4	7.24	5.68	22.47
		3.83	2.65	12.45	48.1	0.97	0.26	2.76	51.9	6.35	4.86	21.62
% Mineral Matter in Refuse = 1.08 x 21.62 + 0.55 x 6.35 = 26.8%. % Coal = 73.2%												
GROUP B												
PENNA	1.40	2.63	2.09	11.15	80.4	0.93	0.28	4.74	19.6	9.62	9.53	37.43
OHIO	1.40	2.45	1.74	10.98	81.4	0.92	0.24	3.93	18.6	9.17	8.28	[illegible].83
OHIO	1.40	2.67	1.72	7.70	85.8	0.93	0.26	2.89	14.2	13.22	10.55	36.79
PENNA	1.40	2.25	1.44	12.60	73.3	0.94	0.24	4.93	26.7	4.66	4.74	33.64
PENNA	1.40	2.47	1.60	12.80	79.8	0.88	0.20	4.89	20.2	8.75	7.12	44.03
		2.49	1.72	11.04	80.1	0.92	0.24	4.28	19.9	9.08	7.64	38.74
% Mineral Matter in Refuse = 1.08 x 38.74 + 0.55 x 9.08 = 46.8%. % Coal = 53.2%												
GROUP C												
OHIO	1.90	1.44	0.80	9.38	95.7	0.78	0.24	6.52	4.3	16.22	13.14	72.99
OHIO	1.90	2.42	1.67	11.02	91.8	0.59	0.17	5.67	8.2	22.95	18.46	70.95
OHIO	1.90	2.52	1.70	9.88	93.2	1.04	0.45	5.82	6.8	22.72	18.32	65.48
PENNA	1.90	2.65	2.09	15.45	87.4	0.67	0.23	7.68	12.6	16.39	15.03	69.32
PENNA	1.90	2.64	1.79	15.56	86.4	1.06	0.38	6.42	13.6	12.68	10.78	73.67
W. VA.	1.90	1.59	0.99	13.52	92.1	0.76	0.26	8.16	7.9	11.28	9.44	76.08
		2.21	1.51	12.47	91.1	0.81	0.30	6.71	8.9	17.04	14.20	71.41
% Mineral Matter in Refuse = 1.08 x 71.41 + 0.55 x 17.04 = 86.5%. % Coal = 13.5%												
	1.40				87.0	0.72	0.17	4.50	13.0	6.26	4.94	42.04
	1.40				83.9	0.52	0.09	3.81	16.1	12.34	9.88	48.59
	1.40				83.1	0.75	0.24	3.37	16.9	11.20	7.69	41.87
	1.40				71.8	0.57	0.12	4.89	28.2	7.94	7.12	42.32
	1.40				76.1	0.88	0.23	4.26	23.9	8.27	6.77	51.56
	1.40				79.7	0.67	0.16	5.57	20.3	5.21	4.22	44.74
					80.3	0.68	0.17	4.40	19.7	8.54	6.77	45.19
% Mineral Matter in Refuse = 1.08 x 45.19 + 0.55 x 8.54 = 53.5%. % Coal = 46.5%												
GROUP D												
ILL.	1.90	1.15	0.60	15.12	91.0	0.95	0.35	8.34	9.1	3.12	3.07	82.72
PENNA	1.60	1.21	0.71	24.71	70.3	0.66	0.14	7.92	29.7	2.50	2.07	64.44
PENNA	1.90	1.50	0.85	12.76	91.7	0.75	0.18	8.00	8.3	9.82	8.29	65.32
PENNA	1.90	1.71	1.20	15.21	88.0	0.93	0.40	7.43	12.0	7.44	7.08	72.26
PENNA	1.60	1.87	1.41	19.47	75.5	0.79	0.35	8.94	24.5	5.22	4.68	68.83
PENNA	1.90	2.34	1.74	20.17	84.0	0.96	0.33	9.41	16.0	9.59	9.15	76.65
		1.63	1.08	17.91	83.4	0.84	0.29	8.34	16.6	6.28	5.72	70.87
% Mineral Matter in Refuse = 1.08 x 70.87 + 0.55 x 6.28 = 80.0%. % Coal = 20.0%												
	1.40				72.3	0.91	0.28	4.80	27.7	1.76	1.42	42.07
	1.40				57.2	0.64	0.10	4.79	42.8	1.97	1.53	51.33
	1.40				74.2	0.70	0.09	3.61	25.8	3.79	3.03	39.07
	1.40				77.3	0.81	0.27	4.99	22.7	4.78	4.38	49.99
	1.40				58.4	0.68	0.24	5.21	41.6	3.54	3.06	46.51
	1.40				63.6	0.94	0.20	4.76	36.4	4.97	4.43	47.09
					67.1	0.76	0.20	4.69	32.9	3.47	2.98	46.01
% Mineral Matter in Refuse = 1.08 x 46.01 + 0.55 x 3.47 = 51.6%. % Coal = 48.4%												
GROUP E												
PENNA	1.60	1.18	0.78	33.70	59.3	0.66	0.15	8.25	40.7	1.94	1.70	70.79
PENNA	1.40	3.58	2.84	21.48	68.0	1.00	0.44	4.33	32.0	9.07	7.94	57.93
W. VA.	1.40	2.42	1.83	33.58	37.8	1.00	0.30	6.41	62.2	3.28	2.76	50.09
W. VA.	1.40	2.99	2.35	27.70	47.6	0.76	0.28	6.14	52.4	5.01	4.24	47.28
PENNA	1.60	1.78	1.17	27.92	66.5	0.89	0.18	8.44	33.5	3.54	3.15	66.60
		2.39	1.79	28.88	55.8	0.86	0.23	6.71	44.2	4.57	3.96	58.54
% Mineral Matter in Refuse = 1.08 x 58.54 + 0.55 x 4.57 = 65.7%. % Coal = 34.3%												

S_t = Total sulfur S_p = Pyritic sulfur   Parr Formula: % Mineral Matter = 1.08 x % ash + 0.55 x % S_t

The three refuse analyses represented by groups A, B, and C were used in the process calculations. The total material matter in the refuse was calculated by the Parr formula and also appears in the table.

Appendix B

BASIS FOR ELECTRIC UTILITY LOAD REQUIREMENTS
(Base Power Plant)

Load		Case 1	Case 2
Net capacity (Mw)		750	1000
Capacity factor (percent)		75	75
Annual output (billion kwh)		4.93	6.57
Capital Cost			
Present day cost	(millions $)	81.6	99.6
Escalation	(millions $)	0	0
Escalated cost	(millions $)	81.6	99.6
Owner's cost	(millions $)		
Interest during construction		7.3	9.0
Other		3.3	4.0
Total cost	(millions $)	92.2	112.6
	($/kw)	122.9	112.6
Fuel Cost Determination			
Net heat rate (Btu/kwh)		9000	9000
Initial fuel cost (cents/million Btu)		19.80	19.80
Fuel cost escalation (percent/yr)		1.50	1.50
Levelized fuel cost (cents/million Btu)		22.06	22.06
Duration of fuel study (yrs)		20	20
Levelized Annual Cost			
Fixed charges	(millions $)	13.0	15.9
Operating and maintenance	(millions $)	1.5	1.8
Fuel	(millions $)	9.8	13.1
Total annual cost	(millions $)	24.3	30.8
	(mills/kwh)	4.94	4.69

Basis for Annual Fixed Charge Determination	(Same for both Case 1 and 2)
Book life (yrs)	30
Tax life (yrs)	30
Rate of return on investment (percent)	8.0
Interest rate (percent)	6.0
Income tax rate (percent)[1]	55.5
Other tax rate (percent)	2.0
Debt financing (percent)	50.0
Income tax, credit (percent)	3.0
Book depreciation	Straight line
Tax depreciation	Double declining balance
Duration of study (yrs)	30

1. includes Federal and State income tax

Appendix C

BASIS FOR POWER PLANTS WHICH SUPPLEMENT REFUSE-FIRED POWER PLANTS

(Alternate Power Plant)

Load	Case 1	Case 2	Case 3	Case 4
Net capacity (Mw)	250	450	500	550
Capacity factor (percent)	55	68	65	71
Annual output (billion kwh)	1.20	2.70	2.85	3.44
Capital Cost[1]				
Total cost (millions $)	40.4	62.8	66.9	73.1
($/kw)	161.8	139.7	133.9	132.8
Fuel Cost Determination				
Net heat rate (Btu/kwh)	9350	9100	9000	9000
Initial fuel cost (cents/million Btu)	19.8	19.8	19.8	19.8
Fuel cost escalation (percent/yr)	1.5	1.5	1.5	1.5
Levelized fuel cost (cents/ million Btu)	22.06	22.06	22.06	22.06
Duration of fuel study (yrs)	20	20	20	20
Levelized Annual Cost				
Fixed charges (millions $)	5.7	8.9	9.5	10.3
Operating and maintenance (millions $)	0.7	1.0	1.1	1.2
Fuel (millions $)	2.5	5.4	5.6	6.8
Total annual cost (millions $)	8.9	15.3	16.2	18.3
(mills/kwh)	7.37	5.69	5.70	5.34

Basis For Annual Fixed Charge Determination

Same as for Base Power Plant - See Appendix B

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[1] Capital costs determined by scaling from Base Plant capital costs (Appendix B). A scale factor of 0.75 was used in all cases.

Appendix D

BASIS FOR COAL-REFUSE-FIRED

(Refuse Power Pla
Estimate Order of Mag

Part I. Capital Cost[1] for Cyclone and Pulverized-Fuel Power Plants

		200	200
Steam generator design (P = pulverized fuel C = cyclone)		P	C
Steam generator[2]	(thousands $)	9,720	9,720
Draft system	(thousands $)	750	750
Coal handling system	(thousands $)	1,730	1,730
Ash handling system[3]	(thousands $)	1,860	1,360
Other	(thousands $)	23,940	23,940
Total	(thousands $)	38,000	37,500

1. Capital costs include: all direct and indirect field costs, engineering, c
interest during construction, and other owner ex
All costs based on present-day (late 1967) costs.
2. In case of pulverized fuel plants includes pulverizers.
3. Includes precipitator necessary for particulate control in a refuse-fired
is not equipped for sulfur oxides removal.

Part II. Capital Cost Used to Evaluate Refuse Power Plants

			Net Capacity (M
Load		200	300
Capacity factor	(percent)	85	85
Annual output	(billion kwh)	1.49	2.24
Capital Cost			
Present day cost	(millions $)	33.4	45.2
Escalation	(millions $)	0	0
Escalated cost	(millions $)	33.4	45.2
Owner's cost	(millions $)		
Interest during construction		3.0	4.1
Other		1.4	1.8
Total cost	(millions $)	37.8	51.1
	($/kw)	188.8	170.3

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Appendix D

-REFUSE-FIRED POWER PLANTS

efuse Power Plant)
ate Order of Magnitude

Plants

Net Capacity (Mw)				
200	300	300	500	500
C	P	C	P	C
9,720	13,160	13,160	19,160	19,160
750	1,010	1,010	1,470	1,470
1,730	2,350	2,350	3,420	3,420
1,360	2,620	1,840	3,720	1,690
23,940	32,360	32,360	47,130	47,130
37,500	51,500	50,720	74,900	73,870

s, engineering, contingency,
d other owner expenditures.
(late 1967) costs.

in a refuse-fired plant which

Net Capacity (Mw)	
300	500
85	85
2.24	3.72
45.2	65.8
0	0
45.2	65.8
4.1	5.9
1.8	2.6
51.1	74.3
170.3	148.8

B

Appendix D (Cont'd.)

Part III. Basis for Annual Fixed Charge Determination

Book life	(yrs.)	30
Tax life	(yrs.)	30
Rate of return on investment	(percent)	8.0
Interest rate	(percent)	6.0
Income tax rate[4]	(percent)	55.5
Other tax rate	(percent)	2.0
Debt financing	(percent)	50.0
Income tax credit	(percent)	3.0
Book depreciation		Straight line
Tax depreciation		Double declining balance
Duration of study	(yrs.)	30

[4] Includes Federal and State income tax.

Part IV. Annual, Levelized Operating and Maintenance Costs Used for Refuse Power Plants

		Net Capacity (Mw)		
		200	300	500
Capacity factor	(percent)	85	85	85
Annual operating and maintenance cost	(millions $)	0.7	0.9	1.4

Part V. Refuse Energy Value Determination[5]

					Net C
		200	200	200	300
Capacity factor	(percent)	85	83	80	85
Refuse ash concentration	(percent, wet)	25.5	34.0	42.5	25.5
Steam generator efficiency	(percent)	85	85	85	83
Net heat rate	(Btu/kwh)	9,680	9,920	10,300	9,420
Refuse cost escalation	(percent/yr)	1.5	1.5	1.5	1.5
Duration of fuel study	(yrs.)	20	20	20	20

[5] For higher heating values, see Appendix

x D (Cont'd.)

y (Mw)

500
85
1.4

Net Capacity (Mw)

200	300	300	300	500	500	500
80	85	83	80	85	83	80
42.5	25.5	34.0	42.5	25.5	34.0	42.5
85	83	83	83	80	80	80
10,300	9,420	9,650	10,000	9,320	9,550	9,900
1.5	1.5	1.5	1.5	1.5	1.5	1.5
20	20	20	20	20	20	20

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Part VI. Refuse Quantities Required by Refuse Power Plants

Refuse Sulfur Concentration (percent, wet)	Refuse Power Plant →	200 Mw		
	Refuse Ash Concentration (percent, wet) →	25.5	34.0	42.5
		(short tons/hour)		
4.25		—	301	—
6.38		253	304	381
8.50		255	307	385
10.62		—	309	389
12.75		—	—	393

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ppendix D (Cont'd.)

200 Mw		300 Mw			500 Mw		
34.0	42.5	25.5	34.0	42.5	25.5	34.0	42.5
t tons/hour)		(short tons/hour)			(short tons/hour)		
301	—	—	436	—	—	702	—
304	381	366	440	550	589	708	886
307	385	369	444	557	594	714	896
309	389	—	447	563	—	720	906
—	393	—	—	569	—	—	916

B

Appendix E

SULFUR RECOVERY PLANT INVESTMENT

Part I - Estimates For Fixed and Working Capital Costs

	Refuse Power Plant Net Capacity (Mw)		
	200	300	500
Fixed capital cost[1] (millions $)	8.3	11.4	17.2
Working capital[2] (millions $)	0.4	0.6	0.9
Total (millions $)	8.7	12.0	18.1
($/kw)	43.5	40.2	36.2

1. Fixed capital cost estimates based on exponential scaling from the fixed capital cost estimated for a 1000 Mw conventional, coal-fired power plant. This capital cost was estimated at $30 million. A scale factor of 0.80 was used.
2. Estimated to be 15 percent of fixed capital cost.

Part II - Financing Basis

Debt financing (percent of total capital cost)	30.0
Interest on loan (percent/yr)	6.0
Loan life (years)	10.0
Interest received on unused portion of loan (percent/yr)	5.0
Sinking fund interest rate (percent/yr)	5.0

Part III - Estimates of Total Investment and Equity Investment

Fixed and working capital (millions $)	8.7	12.0	18.1
Interest during construction (millions $)	0.5	0.7	1.0
Total investment (millions $)	9.2	12.7	19.1
($/kw)	45.9	42.3	38.2
Loan (millions $)	2.6	3.6	5.4
Equity investment (millions $)	6.5	9.0	13.6
($/kw)	32.6	30.1	27.2

Part IV - Discounted Equity Investment

Discounted rate of return (percent)	12.5	12.5	12.5
Years of construction (years)	3.0	3.0	3.0
Discounted equity investment (millions $)	5.1	7.1	10.6

Appendix F

SULFUR RECOVERY PLANT SULFURIC ACID PRODUCTION

(H_2SO_4 concentration = 77.7 wt. percent)
Production in short tons/day

Refuse Ash Concentration (percent, wt.)	Refuse Sulfur Concentration (percent, wet)	Refuse Power Plant Net Capacity (Mw)		
		200	300	500
25.5	6.38	532	770	1,240
25.5	8.50	714	1,033	1,664
34.0	4.25	422	611	984
34.0	6.38	640	925	1,490
34.0	8.50	860	1,243	2,002
34.0	10.62	1,083	1,567	2,523
42.5	6.38	801	1,158	1,865
42.5	8.50	1,079	1,561	2,513
42.5	10.62	1,362	1,971	3,173
42.5	12.75	1,653	2,392	3,851

Appendix G

BASIS SULFUR RECOVERY PLANT O

Part I. Calculated Combustion Gas Volumes

Refuse Ash Concentration (percent, wet)	Refuse Sulfur Concentration (percent, wet)	200 Mw Refuse Power Plant Precipitator[1] (acfm)	I. D. Fan[2] (acfm)	300 M Preci (a
25.5	6.38	1,160,650	618,450	1,6
25.5	8.50	1,158,050	617,100	1,6
34.0	4.25	1,174,700	625,950	1,6
34.0	6.38	1,171,550	624,300	1,6
34.0	8.50	1,169,600	623,250	1,6
34.0	10.62	1,166,550	621,600	1,6
42.5	6.38	1,189,300	633,750	1,7
42.5	8.50	1,196,700	637,650	1,7
42.5	10.62	1,183,050	630,400	1,7
42.5	12.75	1,179,400	628,450	1,7

1. Actual volume of gas to electrostatic precipitator.
2. Actual volume of gas to induced draft fans.

Part II. Basis for Power Consumption

A. Induced Draft Fans

Gas volume - from Part I
Head - 25 inches H_2O
Static efficiency - 70 percent
Motor efficiency - 95 percent

B. Electrostatic Precipitator

Gas volume - from Pa
Efficiency - 99.9 per
Power requirement - 800 watt

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Appendix G

VERY PLANT OPERATING COSTS

ant n[2])	300 Mw Refuse Power Plant		500 Mw Refuse Power Plant	
	Precipitator[1] (acfm)	I. D. Fan[2] (acfm)	Precipitator[1] (acfm)	I. D. Fan[2] (acfm)
	1,678,750	894,550	2,703,000	1,440,300
	1,675,000	892,550	2,696,950	1,437,100
	1,699,100	905,350	2,735,710	1,457,750
	1,694,550	902,950	2,728,400	1,453,850
	1,691,700	901,450	2,723,800	1,451,400
	1,687,250	899,100	2,716,700	1,447,600
	1,720,200	916,600	2,769,700	1,475,850
	1,730,900	922,300	2,786,900	1,485,000
	1,711,100	911,800	2,755,100	1,468,050
	1,705,900	909,000	2,746,650	1,463,550

recipitator

- from Part I
- 99.9 percent

nent - 800 watts/1000 acfm

C. Other Power Consumption

Estimated at 10 percent of induced draft fan and electrostatic precipitator consumption.

B

Part III. Catalyst Requirements

Reaction rate - Calder bank relation (CEP, vol. 49, no. 11, Nov. 1953)
Equilibrium constant - K = 103 (Duecker, W. W., West, J. R., The Manufacture of Sulfuric Acid, 1959)
Catalyst bed temperature - 455° C (assumed constant)
Total pressure - 1 atmosphere
Conversion - 90 percent
Life - 2 years

Part. IV. Cooling Water Consumption

Estimated at 50 gal. /ton of sulfuric acid

Part V. Labor Requirement

Operating labor - 48 manhours/day

Part VI. Unit Costs

Power - 8 mills/kwh
Catalyst - $1.40/liter
Cooling water - $0.10/1000 gal.
Supplies - 0.20 percent/year of capital cost
Chemicals and laboratory - $5000/year
Operating labor - $4.00/hour
Supervision - 15 percent of operating labor
Payroll extras - 25 percent of labor plus supervision
Maintenance (labor and materials) - 6.0 percent/year of capital cost
General overhead - 30 percent of payroll

Appendix H

BASIS SULFUR RECOVERY PLANT CASH FLOW CALCULATIONS

Federal income tax rate	48 percent
State income tax rate	7 percent
Other taxes	2 percent of fixed capital cost
Investment tax credit (taken in first year)	7 percent
Plant tax life	20 years
Depreciation method	double declining balance with switch to straight line
Plant operating life	20 years

Appendix I

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